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Western Canada Coal Liquefaction Study
Prefeasibility Study

Prepared for:

British Columbia Resources Investment Corporation
Petro-Canada Exploration Incorporated
Westcoast Transmission Company Limited
British Columbia Coal Limited
Energy, Mines and Resources Canada

Volume I
Technical Evaluation and Recommendations

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Kilborn and Badger have followed standard professional procedures in preparing the contents of this Report. It is based in part on details, information and assumptions provided by others and Kilborn and Badger, therefore, cannot guarantee the correctness of the Report; but to the extent of their investigation and within the scope of the work delegated to them, Kilborn and Badger believe the Report is substantially correct.

1.0 INTRODUCTION

1.1 PROJECT DESCRIPTION

British Columbia has the second largest coal reserves of any province in Canada. Many of these reserves are located in the Rocky Mountain region adjacent to the eastern border of the province. Although much of this bituminous low sulphur coal is mineable by relatively cheap open pit methods development is currently limited by demand. Local coal markets are not expanding and the location of these reserves is very distant from the larger coal markets in Pacific Rim countries and Eastern Canada. Although development of some of these deposits in south eastern and north eastern British Columbia has recently begun, mainly for export to Pacific Rim countries, other large deposits, including some owned by Petro Canada, remain undeveloped.

Since British Columbia depends on extra-provincial supply sources for approximately 75% of its current oil requirements, the potential for liquefying these coals into petroleum substitutes deserves investigation. Coal conversion into liquid transportation fuels, using advanced liquefaction technology, could offer the opportunity for savings on oil imports, movement towards Canadian and British Columbian oil self-sufficiency and long term coal utilization. In recognition of this potential the Western Canada Coal Liquefaction Consortium was created to initiate the development of a coal-based synthetic liquid fuels industry in Western Canada.

In order to evaluate the potential of liquefying bituminous coals from the Monkman and Sparwood areas of British Columbia, the consortium retained Kilborn Limited to conduct a Prefeasibility Study. The primary objective of Kilborn's work was to assess the technical and economic feasibility of liquefying the coals in these locations. Kilborn retained Badger Energy Inc. of Cambridge, Massachusetts as a sub-consultant to provide assistance for this study.

This study is part of the ongoing oil substitution program in Canada involving the Federal Government, Provincial Governments and private industry. Public support for the study was made available through the Canada Centre for Mineral and Energy Technology (CANMET). The members of the consortium for the Western Canada Coal Liquefaction Study are:

The British Columbia Resources Investment Corporation
Petro-Canada Exploration Inc.
Westcoast Transmission Company Limited.

1.2 SCOPE OF PRE-FEASIBILITY STUDY

The principal original objective of the Prefeasibility Study was to evaluate the technical and economic feasibility of conversion of bituminous coals in Eastern British Columbia to transportation liquids using direct liquefaction technology. The coals and processes to be considered in the study were selected by the Western Canada Coal Liquefaction Study Consortium. These selections, made prior to commencement of this study, were as follows:

Nominated Coals	Monkman Sparwood
Nominated Processes	SRC II H-Coal Exxon Donor Solvent NCB/BP Solvent Extraction Ruhrkohle/Veba

The technical and economic merits of each process were to be evaluated in liquefying both the nominated coals. It was originally intended to obtain test data specific to these coals from each of the process licensors. This data would then be developed to permit the optimum choice of process and coal in order for a technical economic and environmental assessment to be made of a commercial installation.

Very early in the study it was decided to incorporate coal testing by Canadian laboratories into the program to maximize Canadian participation while minimizing testing by individual process developers. In this manner, not only was experience on coal liquefaction technology gained in Canada, but testing costs were minimized as duplicate tests were not conducted by several process developers.

Test data developed by B.C. Research and Sandwell Beak Research Group (SBRG) were provided to the process developers as an aid in the evaluation of the potential of their processes to liquefy the nominated B.C. Coals. Limited testing was conducted by NCB and HRI to better assess the capability of the NCB/BP LSE and H-Coal processes, resepctively, to liquefy the high rank B.C. Coals. Their data, or the results thereof, are incorporated in the report. A test program, specific to the German process, is currently being conducted by Veba Oel; results will not be available in time, however, to be incorporated into this report.

As the test programs conducted by B.C. Research, SBRG and HRI and relevant literature information all predicted low distillate yields for the nominated B.C. Coals, it was decided to develop preliminary costs and economics for a 'typical' direct coal liquefaction process, utilizing Greenhills coal. It was further decided to evaluate the coal liquefaction process on the basis of the production of a coal-derived synthetic crude product, not finished transportation fuels as was the original concept. Specifications were established for the syncrude fractions (naphtha, middle distillate and heavy distillate) such that the syncrude product could be processed by other Canadian refineries to finished products -- transportation fuels and/or industrial fuels and home heating oil. Results of this technical and economic assessment would be used to decide the next phase in the development of the coal liquefaction project, should further development be warranted.

Specific tasks which were undertaken included:

- A literature search to obtain the maximum quantity of information relating to the nominated processes and coals.

- Evaluation of this literature information.

- Arranging microautoclave and continuous testing of the nominated coals at B.C. Research and Sandwell Beak Research Group laboratories to evaluate reactivity of the nominated coals under process conditions similar to those used by the process developers.

- Review and evaluation of the liquefaction potential of the nominated coals based on data obtained from the tests described above.

- Contact with the process licensors to obtain additional data and to seek their participation to meet the study requirements.

- Visits to specific process licensors research laboratories to evaluate possible predictions from small scale testing facilities and to evaluate their predictions for the nominated coals.

- Review of data provided by process licensors at presentations to the Western Canada Coal Liquefaction Study Consortium and as a result of their own test programs.

- A technical and economic assessment of a 'typical' direct liquefaction process, feeding 10,000 tonne/day of washed Greenhills coal and producing synthetic crude.

- A preliminary socio-economic and environmental overview of the expected impacts from construction and operation of the liquefaction facility.

- Preparation of preliminary capital and operating cost estimates for the 'typical' liquefaction facility.

- Preparation of preliminary economic sensitivity analyses for the liquefaction facility including sensitivity to the price of coal.

- Preparation of preliminary comparative economics for alternative processing configurations to the 'typical' liquefaction facility, selected to improve project profitability.

- Preparation of a comprehesive report summarizing the results of the Prefeasibility Study.

The technical information relating to the above tasks are presented in Volumes I and III, and the economic data are presented in Volume II of this report.

2.0 CONCLUSIONS AND RECOMMENDATIONS

The conclusions and recommendations resulting from this prefeasibility study are highlighted herein.

2.1 CONCLUSIONS

2.1.1 Coal Reactivity

Both the B.C. coals from the Greenhills and Monkman reserves are high rank Cretacious coals which have undergone severe tectonic deformation. The high inert content and low volatile matter in these coals render them resistant to high conversion to liquid distillate products via conventional direct liquefaction processes. Little data are available on the conversion of high rank coals as all process developers are concentrating on developing coal liquefaction processes capable of processing lower rank coals, usually high volatile "B" and "C" bituminous coals, sub-bituminous coal and lignite. Only the Germans have substantial data on processing high volatile "A" bituminous coal. By comparison, the nominated B.C. coals are medium volatile or low volatile bituminous coals.

The results of the B.C. Research microautoclave tests confirmed that the B.C. coals are difficult to liquefy. Distillate yields were only 42-50 percent under optimum test conditions, far less than Illinois No. 6 coal which was consistently 10-30 percent more reactive. Illinois No. 6 coal, a high volatile "B" bituminous coal, is the reference coal used by most process developers in the United States. Throughout the tests, however, the Greenhills coal demonstrated a marginally better potential for liquefaction than the Monkman coal, consistently showing a 1-5 percent better distillate yield than the Monkman coal.

In parallel tests conducted by B.C. Research, 'simulation' of the German process with Red Mud catalyst demonstrated 10-20 percent better yield than with the Hydrocarbon Research H-Coal Process 'simulated' tests with HDS-1442A catalyst. However, the HDS-1442A tests were performed with hydrogenated anthracene oil as solvent, whereas the Red Mud catalyst tests were conducted with coal-derived solvent provided by Bergbau-Forschung. Tests with the same catalyst and the different solvents showed a significant dependence upon the nature of the solvent. From the B.C. Research tests it was not possible to determine which catalyst was favoured -- HDS-1442A as used in the H-Coal process or Red Mud (iron oxide) as used in the German process. The tests did show, however, that a non-catalytic process, such as Exxon Donor Solvent (EDS) or Solvent Refined Coal International Inc. (SRC-II), is definitely not favoured. No data were obtained relative to the National Coal Board/British Petroleum (NCB/BP) process.

Results of a subsequent HRI microautoclave test program, designed to simulate coal-derived solvent and vacuum bottoms recycle operation, confirmed that the B.C. coals are difficult to liquefy. Predicted H-Coal distillate yields were 43 percent, maf coal basis, for Greenhills coal and several percent less for Monkman coal. Of the predicted distillate yield, the greatest fraction was heavy distillate (345- 525^{o}C) with lesser amounts of middle distillate (205-345^{o}C) and naphtha (C_5-205^{o}C). By comparison, distillate oil yields with Illinois No. 6 coal in the Syncrude mode are 48-50 percent, maf coal basis. Raw product fractions typically consist of 30-34 percent C_5+ naphtha, 40-44 percent middle distillate and 22-28 percent heavy distillate, depending on reaction conditions and catalyst activity.

The continuous tests, performed by Sandwell Beak Research Group (SBRG), showed relatively high conversion for both coals but the distillate yields were low. A minimum and maximum range of distillate yields of 19-34 percent was predicted for the once-through solvent operation, using hydrogenated anthracene as solvent. No significant difference in distillate yields was observed in the two tests -- Greenhills coal with Red Mud catalyst and Monkman coals with HDS-1442A catalyst -- at least partially because equipment limitations prevented the Greenhills coal/Red Mud catalyst test from being conducted at the optimum temperature identified in the B.C. Research program.

2.1.2 Process Selection

Of the five direct liquefaction processes considered, SRC-II and EDS were eliminated on the basis of the results of the B.C. Research program and other pertinent information provided by the process developers. The favourability of the catalytic processes in the microautoclave tests, a demonstrated correlation between distillate yield and volatile matter for the EDS process, in particular, and the uncertainty of the commercialization plans of the two developers influenced the elimination of these processes from this study. SRC-II appear to have no commitment for commercialization of their process since cancellation of the 5400 tonne per day demonstration plant project in 1981.

Screening of any of the three remaining processes -- H-Coal, Ruhrkohle/Veba and NCB/BP Liquid Solvent Extraction -- was not possible, based on the results of the test programs and process developer predictions. Although a preliminary prediction of distillate yields of 54-56 percent made by NCB for their NCB/BP process appears very attractive, only limited confidence can be placed in the results as the process is in the early stages of development and very little data are available on integrated liquid solvent extraction and hydrocracking steps of the process with high rank (medium or low volatile bituminous) coals. Successful operation at the pilot plant scale, 25 tonne per day, must be demonstrated before the process can be given serious consideration.

Data available at the present time does not permit a selection to be made between the H-Coal and Ruhrkohle/Veba direct liquefaction processes. Once the Veba Oel autoclave test program is completed, additional data will be forthcoming, such that it should be possible to rank the two processes according to their suitability to process the B.C. coals, Greenhills or Monkman. The maximum distillate oil yield predicted by HR1 for the H-Coal process with vacuum bottoms was 43 percent.

In both the H-Coal and Ruhrkohle/Veba processes, the raw liquid products require substantial upgrading to obtain transportation fuels as they contain high concentrations of heteroatoms, particularly nitrogen and oxygen, and are deficient in hydrogen. In comparison the NCB/BP process produces liquid products that require very little upgrading to meet transportation fuel specifications. These factors will become important in final process selection.

2.1.3 Process Evaluation

To assist the Western Canada Coal Liquefaction Study Consortium in deciding whether or not further testing and evaluation of the three remaining liquefaction processes is warranted, a preliminary financial evaluation of a 'typical' coal liquefaction process was prepared. A preliminary process evaluation was completed and order-of-magnitude capital and operating cost estimates were prepared for a 'typical' coal liquefaction plant designed to process 10 000 tonne per day beneficiated coal and yield an assumed 40 percent liquid distillates (C_5 - 525°C), or a nominal 24 000 barrels per day of distillate products.

The H-Coal process was chosen to represent the 'typical' coal liquefaction process and Greenhills coal was selected on the basis of its marginally better distillate yield in the microautoclave test programs. In no way does this evaluation of the H-Coal process signify that it is better suited for the liquefaction of the B.C. coals than the other two remaining processes, Ruhrkohle/Veba and NCB/BP. It was selected as representing all three processes for this preliminary evaluation, simply on the basis of the quality of information available to Kilborn and Badger. Similarly, use of Greenhills coal does not preclude the future development of a coal liquefaction project with Monkman coal.

The 40 percent distillate yield was chosen on the basis of the data from the B.C. Research, SBRG and HRI test programs as being representative of the conventional configuration of the H-Coal process feeding Greenhills coal and on the assumption that 40 percent would be the minimum possible distillate yield that could give a positive economic analysis. The process evaluation was conducted on the basis of processing Greenhills coal with the exception of the assumed division of the distillate product into naphtha, middle distillate and heavy distillate fractions and the assumed analyses of the raw liquid fractions. The split between the three liquid fractions was prorated from HRI data on Illinios No. 6 coal to give a 40 percent distillate yield and the elemental analyses of the raw products were assumed to be the same as for the Illinois No. 6 coal. Similarly, the yield of light hydrocarbon gases (C_1 - C_4) was assumed to be the same as with Illinois No. 6 coal.

By using this basis for the process evaluation, material balances reflect hydrogen consumption, sulphur production and ammonia production levels for Greenhills coal to synthetic crude. It is recognized that the assumed division of the distillate products into the three fractions resulted in higher naphtha and middle distillate contents than demonstrated by the test programs, as shown below:

	Raw Product Fraction, Percentage	
	Assumed	HRI Test Results
Naphtha	28	4 - 10
Middle Distillate	40	14 - 46
Heavy Distillate	32	82 - 44

This difference has been taken into account in the financial analysis as various discounts were applied to the syncrude price and a sensitivity analysis on the effect of product revenue (in reality, syncrude prices) on project rate of return was performed.

Main products produced by the 'typical' liquefaction facility include:

Propane	3725 barrels/day
Butane	2940 barrels/day
Naphtha Fraction	9045 barrels/day
Middle Distillate Fraction	11020 barrels/day
Heavy Distillate Fraction	4215 barrels/day
Sulphur	35 tonne/day
Ammonia	116 tonne/day

Hydrotreating facilities were provided, based on UOP information, to upgrade the raw coal-derived products to high grade petroleum fractions that could be sold directly as industrial fuels or upgraded in a Canadian refinery to transportation fuels. The estimated capital cost associated with hydrotreating (hydrotreaters, hydrogen production, associated utility and offsite facilities) is in the order of $275 million. Direct operating costs for the hydrotreating facilities is in the order of $2.50 - $3.00 per barrel of hydrotreated product; if capital charges are included at 15 percent straight line, total upgrading cost becomes $8.00 - $8.50 per barrel.

Total hydrogen requirements for the coal liquefaction plant are 262 000 Nm^3 per hour or 235 million SCFD, 206 000 Nm^3 per hour for the liquefaction section and 56 000 Nm^3 per hour for hydrotreating.

The estimated capital cost for the Base Case coal liquefaction facility is $2.02 billion in 1982 Canadian dollars. Annual operating costs, estimated for two coal prices, and annual gross revenues, estimated on the basis of the September 1981 Oil Pricing Agreement and the April 13, 1982 Announcement by the Province of Alberta, are given below in 1982 Canadian dollars. Discounted cash flow rates of return to 100 percent equity are included for each case.

	$80/tonne coal	$50/tonne coal
September 1981 Agreement		
Annual Revenue, $1000	437 500	437 500
Operating Cost, $1000	393 080	300 080
Gross Margin, $1000	44 420	137 420
DCF Rate of Return, %	4.6	10.2
April 13, 1982 Announcement		
Annual Revenue, $1000	349 630	349 630
Operating Cost, $1000	393 080	300 080
Gross Margin, $1000	Negative	49 550
DCF Rate of Return, %	Negative	4.3

According to the April 13, 1982 Announcement, new oil prices effective July 1986, are forecast to be $60.85 per barrel, compared to $77.40 per barrel per the September 1981 Agreement. This difference in forecast new oil price is reflected in the annual gross revenue estimates shown above and serve to substantiate the extreme sensitivity of the project economics to product prices. Similarly, the data at two coal prices confirm the sensitivity of the project economics to coal prices.

2.1.4 Socio-Economic and Environmental Considerations

The proposed coal liquefaction project will potentially create a number of economic and socio-economic benefits within the East Kootenay region of British Columbia. One of the most significant benefits will be associated with employment requirements generated by the project -- an estimated peak construction workforce of 5000 during a 3-4 year construction period and an estimated permanent operating staff of 1000, both excluding the associated coal mine.

In addition to the direct employment opportunities, the project will generate a number of indirect and induced potential employment opportunities in the service and manufacturing sectors of the economy. These spin-off opportunities and related developments can be influential in stimulating economic activity both within the local community and well beyond. Based on previous experience, the total direct and indirect employment estimated to result from the construction of the facility will be approximately 13 000 within British Columbia. Similarly, during the operational phase the facility will generate an estimated 4000 direct and indirect employment opportunities.

In terms of environmental considerations, it can be anticipated that, whatever impacts this project may have, they will be of relatively lesser degree than might otherwise occur if heavy industrial and mining development were not already underway in the Greenhills area. The emissions anticipated do not appear to be beyond known technologies for their proper handling and control. It must be emphasized, however, that only a cursory environmental overview has been undertaken during this prefeasibility study. Thus, a detailed environmental investigation must be undertaken in the next phase of the project.

2.2 RECOMMENDATIONS

As it was not possible to make a final selection of the process and coal best suited for liquefaction, further testing is required before final selection can be made. The purpose of the technical and economic evaluation of the H-Coal process, as being 'typical' of the H-Coal, Ruhrkohle/Veba and NCB/BP processes, was to determine if further testing is warranted in view of the low distillate yields that were predicted by the microautoclave and continuous bench scale experimental programs.

As the results of the preliminary economic evaluation are discouraging, particularly with reference to the significantly lower New Oil price forecasts contained in the Province of Alberta Announcement of April 13, 1982, further development of the coal liquefaction project should be approached with caution. Further development is warranted to answer major unknowns identified in this prefeasibility study as the possible social benefits associated with the project are as important as, if not more important than, the financial benefits.

The main emphasis of the recommended development program should be placed on the evaluation of process parameters that would make the project more attractive economically, including the following:

- increased liquid distillate yield (above 40 percent) per tonne of feed coal,
- lower coal price and/or alternate source of coal that offers a better potential for liquefaction (lower rank than the Greenhills or Monkman coals) at low cost.

Specifically, it is recommended that a phased development program be undertaken that will ultimately lead to the completion of a detailed feasibility study only if results of the earlier tasks warrant continuation of the project. Tasks to be undertaken in the phased development program include:

- evaluation of possible other B.C. coals which are better suited for liquefaction,
- discussions with leading coal liquefaction process developers (HRI, Veba Oel, NCB/BP) to determine the confidence they have in the ability of their processes to give distillate yields in excess of 40 percent with Greenhills, Monkman or other nominated B.C. coal,
- continuous bench scale testing of one or two leading coal liquefaction processes to demonstrate the process under optimum operating conditions, to generate yield and hydrogen consumption data and to provide raw product samples for subsequent upgrading tests,
- continuous laboratory scale testing of raw product hydrotreating technology by one of the hydrotreater process licensors to determine required operating conditions to meet desired product specifications,
- preparation of a standard design specification package for a Texaco coal liquefaction residue gasification process,
- conceptual process engineering, cost estimating and financial evaluation for a commercial scale coal liquefaction facility.

Details of the various tasks to be undertaken during the recommended development program, along with indicated decision points, follow.

2.2.1 Evaluation of Other Coals

As the economic attractiveness of the proposed coal liquefaction process is extremely sensitive to product revenue (one factor of which is distillate oil yield per tonne of coal) and to the price of coal, it is recommended that alternate B.C. coals be investigated for use as liquefaction plant feed. Key factors to use in this investigation are:

- adequate reserves to support the coal liquefaction project over a 20 year life, approximately 80-100 million tonne for a 10 000 tonne per day facility,
- low price, $50 per tonne or lower for beneficiated coal (6-8 percent moisture, 8 or less percent ash),
- good liquefaction potential, as indicated by the rank of the coal. In general, high volatile "A" bituminous, or lower rank, coals are more reactive than medium or low volatile bituminous coals.

If suitable other B.C. coals are found, a microautoclave program similar to the B.C. Research program conducted as part of this study should be commissioned to determine the reactivity of these coals relative to Greenhills and Monkman coals. Results of this test program would form the basis for the selection of the coal for subsequent testing by the process developers.

2.2.2 Discussions with Process Developers

Discussions should be initiated with the three leading process developers (HRI, Veba Oel, NCB/BP) to better assess the capability of each process to efficiently convert the nominated B.C. coal to prime liquid transportation fuels. Important issues to be addressed during these discussions include:

- the degree of confidence demonstrated by the process developer in the ability of his process to give distillate yields in excess of 40 percent with Greenhills, Monkman or other nominated B.C. coal,
- the extent of experience in processing high rank coals in autoclave, bench scale and pilot scale equipment,
- a demonstrated commitment by the process developer to continue development of his process to at least a large scale continuous pilot plant stage.

Distillate product yields projected by NCB for the NCB/BP LSE process are substantially above 40 percent and are attractive provided that they can be confirmed in continuous pilot plant testing. During discussions with NCB/BP, firm confirmation should be obtained that continued development of their process, in the form of construction and operation of the proposed 25 tonne per day Point of Ayr pilot plant, will take place. As the NCB/BP process is in the early stages of development, successful demonstration at the 25 tonne per day scale is essential before serious consideration can be given to the process. Development of this pilot plant is heavily dependent on financial backing by private industry. To date, this financial backing has not been confirmed and design and construction of the pilot plant is being delayed.

DECISION POINT - Upon completion of the evaluation of alternate B.C. coals and discussions with coal liquefaction process developers, all information will be thoroughly evaluated to determine whether one, two or zero continuous bench scale test programs should be commissioned. To warrant testing, distillate product yields significantly in excess of 40 percent should be expected, with reasonable assurances from the process developers.

If two processes are to be tested, it is recommended that one process be the NCB/BP process and the other process be one of the two direct liquefaction processes, H-Coal or Ruhrkohle/Veba.

2.2.3 Coal Testing

If the decision is made to proceed with coal testing, continuous bench scale testing should be performed to further evaluate one or two coal liquefaction processes. The main objectives of the continous bench scale testing will be:

- to demonstrate the integrated process under optimum operating conditions,

- to generate yield and hydrogen consumption data under 'simulated' commercial scale conditions,
- to provide raw product samples for analysis and subsequent upgrading tests.

The continuous testing should be conducted in an integrated bench scale unit that tests all major steps in the coal liquefaction process simultaneously, including coal slurrying, reaction, product separation, recycle oil preparation and, in the NCB/BP case, filtration and hydrocracking. If process simulation, particularly recycle oil operation, is conducted in a stepwise manner, distillate oil yields and product losses may not be totally representative of continuous operation.

DECISION POINT - At the completion of the continuous bench scale test program, the test data will be thoroughly evaluated to determine whether or not further development of the coal liquefaction project is warranted and which process is best suited to the nominated coal. For the project to proceed, the test data should confirm a high distillate oil yield from the nominated B.C. coal, such that there would be a reasonable chance that overall project economics would be better than given in this prefeasibility study. In addition to distillate oil yield, other economic factors that might be different than those used in this evaluation should be considered in making this decision.

2.2.4 Additional Testing

If the decision is made to continue the development program, continuous laboratory scale testing of raw product hydrotreating should be performed to demonstrate the technology required to successfully upgrade the raw coal-derived products to specification transportation fuels or, alternatively, industrial fuels.

In addition, Texaco Development Corporation should be commissioned to prepare a standard Type-C process package for their residue gasification process, feeding vacuum residue or filtered solids from the selected coal liquefaction process.

Details of this program cannot be defined until the coal liquefaction process has been selected. For the NCB/BP process, only limited raw product upgrading is required as the final step in their process is a hydrocracking step which produces distillate fractions close to transportation fuel specifications. However, raw products produced by the direct liquefaction processes, H-Coal and Ruhrkohle/Veba, require substantial upgrading to meet transportation fuel specifications. Similarly, residues from the two types of processes (the NCB/BP extraction and the direct liquefaction) are markedly different and must be processed in different manners in the residue gasification process.

2.2.5 Technical and Economic Evaluation

A technical and economic evaluation of one coal liquefaction process would be performed only if the results of the continuous coal liquefaction bench scale test program were sufficiently encouraging to warrant continued development. As discussed in Section 2.2.3, a decision would be made at the conclusion of the test program as to whether or not further work would be performed and which process (if two were tested) was best suited to process the nominated B.C. coal.

2.2.5.1 Engineering Evaluation

The study would incorporate design concepts for the various plant components in sufficient detail as required to prepare cost estimates with an accuracy of ±25% for the selected process. The items to be developed and prepared will incorporate:

- a review of process packages and test data from the selected coal liquefaction process developer, the hydrotreater licensor and Texaco for the residue gasification process,
- obtain process data on other process facilities, including coal preparation, light ends recovery, hydrogen recovery and purification, sour gas scrubbing, sulphur plant and air separation plant,

- preparation of conceptual flowsheets for each process unit,
- determination of product slate for the plant,
- preparation of preliminary flow sheets for offsite facilities including product tankage, effluent treatment, flare system, etc.,
- preparation of preliminary flowsheets for utility systems including steam, air, water and plant fuel systems,
- preparation of material and energy balances for the various plant components,
- determination of catalyst and chemical requirements,
- determination of utility requirements including electricity, steam, cooling water, plant air, instrument air, water, natural gas, etc.,
- determination of operating manpower requirements,
- preparation of conceptual plot plans and layout drawings,
- identification and sizing of major items of equipment and preparation of an equipment list,
- preparation of preliminary project execution plans and schedules, including construction manpower requirements.

2.2.5.2 Estimating

Capital and operating cost estimates will be prepared based on cost data provided by the selected licensors and on the drawings and data developed through the Engineering activities outlined above.

Equipment vendors will be contacted for budget prices on major items of equipment for the project. This information together with cost data obtained from the licensors and other in-house data will be incorporated in an overall project estimate with an accuracy of ±25%.

Operating cost estimates for the complete plant, including cost of manpower, utilities, catalysts, chemicals, feedstocks and maintenance will be prepared. These estimates will be based on data provided by the licensors and in-house experience.

These estimates will form the basis for the economic and financial evaluation of the selected process.

2.2.5.3 Financial Feasibility

Close co-operation between the Consortium and the Consultant will be required during the performance of this task. The basic parameters, assumptions and other variables will be jointly established in order to develop an appropriate economic model for this analysis.

Discounted cash flows will be developed with various combinations of financial structures and several probability factors. These will include elements such as equity participation by government, accelerated depreciation, preferential tax treatment, phased investments, preferential freight rates, government development of infrastructures including shipping facilities, services, roads, energy supplies, etc.

The financial analysis will cover the projected life span of the project with jointly agreed upon interest rates, escalation, internal rate of return, etc. This task will include manipulation of variables individually or in combination to permit the generation of sensitivity and risk analyses.

2.2.5.4 Socio-Economic Evaluation

In estimating the social/economic benefits which the proposed complex will have on the region and the province, a number of factors will be investigated and evaluated. These factors include:

- housing and infrastructure capacity
- availability of employment and manpower skills

- an evaluation of the probable sources of non-local workers required for construction and operating phases and characteristics of incoming employees
- available manpower training programs and probability of their adaptation to suit the project requirements
- anticipated competition for skilled labour demand resulting from other major projects
- the expected personal incomes and forecasted spending patterns
- the nature and extent of demands on community services which will be required as a result of the project
- impact on commercial services such as hotels/motels, restaurants and other economic spin-offs
- impact on municipal services such as water and sewerage systems
- effects of the project on transportation modes by road, rail and air
- evaluation of municipal, regional and provincial revenues
- effects of forecasted distribution pattern of construction expenditures and sales revenues.

2.2.5.5 Environmental Overview

Appropriate government and regulatory agencies will be contacted at the earliest possible stage to obtain all stipulations and guidelines which need to be observed for the execution and operation of the project. This task will include a concise overview identifying those impacts which are likely to occur during the construction and operation of the project. This analysis will be based on available information specific to the selected site and will focus on physical and natural components of the environment.

Factors to be considered in evaluating impacts on physical environment include:

- source and quantity of water requirement and its impact on existing water bodies or ground water
- potential for increase in erosion and sedimentation due to site preparation activities
- effect of drawdown, increased discharges and changes in ambient temperature as a result of plant operation
- evaluation of liquid effluents and gaseous emissions and the potential impacts on water and air quality.

Effects on natural environment will include an analysis of terrestial and aquatic ecologies. This analysis will address concerns relative to present species, sensitive habitats, fisheries, communities and vegetation.

2.2.6 Development Program Costs and Schedule

It is not possible to define an expected cost or schedule for the proposed development program at this time as both are highly dependent on the amount of testing performed and on the actual coal liquefaction processes tested.

Expected costs are presented in the following section for specific tasks associated with the proposed development program to assist the Western Canada Coal Liquefaction Study Consortium in deciding the extent of testing to be performed, if further development of the coal liquefaction project is undertaken. These expected costs should be used as a guide only and should not be considered firm for any of the tasks.

An overall schedule of 12-18 months would be expected for completion of the entire proposed development program. Actual timing and length of schedule is, however, largely dependent on the testing schedules and bench scale equipment availability of the process developers, over which the contractor has no control.

Task	Expected Cost $ Can. 1982
Microautoclave testing of alternate B.C. coals to determine their reactivity relative to Greenhills or Monkman coals.	$ 25 000 - 50 000
Continuous bench scale testing by coal liquefaction process developers to demonstrate all steps of the integrated process and to generate yield and hydrogen consumption data	
- HRI H-Coal	$300 000 - 500 000
- Ruhrkohle/Veba	$250 000 - 300 000
- NCB/BP LSE	$ 90 000 - 150 000[1]
Raw liquid product upgrading testing to demonstrate the technology required to upgrade the raw products to specification transportation fuels.	$ 50 000 - 240 000[2]
Preparation of Type-C process package for the Texaco residue gasification process.	$ 50 000 - 60 000
Technical and economic evaluation of the selected coal liquefaction process.	$600 000 - 800 000

NOTES:

1. The lower number given by NCB for testing the NCB/BP LSE process is based on using a very small (0.5 kg/day) bench unit that is not fully integrated as it does not include simultaneous extraction and hydrocracking with solvent recycle. The higher number reflects testing on an integrated 5 kg/day unit.

2. The range of costs for upgrading testing is large as much more severe upgrading is required for raw products from the direct liquefaction processes, H-Coal and Ruhrkohle/Veba, compared to the NCB/BP liquid solvent extraction process.

3.0 COAL PROPERTIES

3.1 GENERAL

The two coals nominated by the Consortium for consideration as feed coal to the liquefaction process are from the Monkman and Greenhills reserves. These two areas are both near the eastern border of British Columbia, but are over 700 km apart. The Monkman reserves are more northerly being located in an area between Prince George and Dawson Creek while the Greenhills reserves are located in the south eastern corner of the province just to the north of Sparwood. Figure 3.1 shows the two locations.

Both the Greenhills and Monkman reserves contain coals of fairly similar petrography. They are both high rank medium to low volatile bituminous coals with low sulphur contents (<0.5%). Detailed properties of the two coals are given in Tables 3.1 and 3.2.

3.1.1 Monkman Coal Reserve Data

The Monkman coal reserves are located within a 75 km stretch of the inner foothills of the Rocky Mountains in north east British Columbia. The general area is 150 km north east of Prince George and 110 km south west of Dawson Creek. The area which covers 38 000 ha varies in elevation and contains an estimated 2.8 billion tonne of coal.

Several blocks of these coal reserves have been explored and the Duke Mountain Block is presently the focus of studies. In the northern part of this block two areas, known as the Duke and Honeymoon areas have recently been explored. Coal quality information has led to an open pit mining scheme being proposed for both areas. These proposed mines would produce a total of 68.9 million tonne

R.O.M. coal at an average strip ratio of 6 m^3 rock per tonne of raw coal. The annual output when in full production would yield 3.29 million tonne of metallurgical coal. Additional reserves in the block suggest the mine life could be extended well beyond the currently projected life of 16 years and so satisfy the 20 year operating life of the liquefaction project.

3.1.2 Greenhills Coal Reserve Data

The Greenhills coal reserves are located approximately 25 km north of Sparwood in the hills between the Elk and Fording Valleys. The B.C. Coal property area in this region covers an area of approximately 100 sq km. The reserves allocated to this projects are 190 million tonne at an average strip ratio of 9 m^3 rock per tonne of raw coal.

A project is currently under development for recovery of this high quality coal and open pit strip mining operations are expected to commence in 1983. This development will produce an annual output of 1.8 million tonne of metallurgical coal with additional quantities of thermal coal with firm plans for increasing up to 3.6 million tonne per year. The reserves allocated to this project are at higher strip ratios (average 9:1 metric).

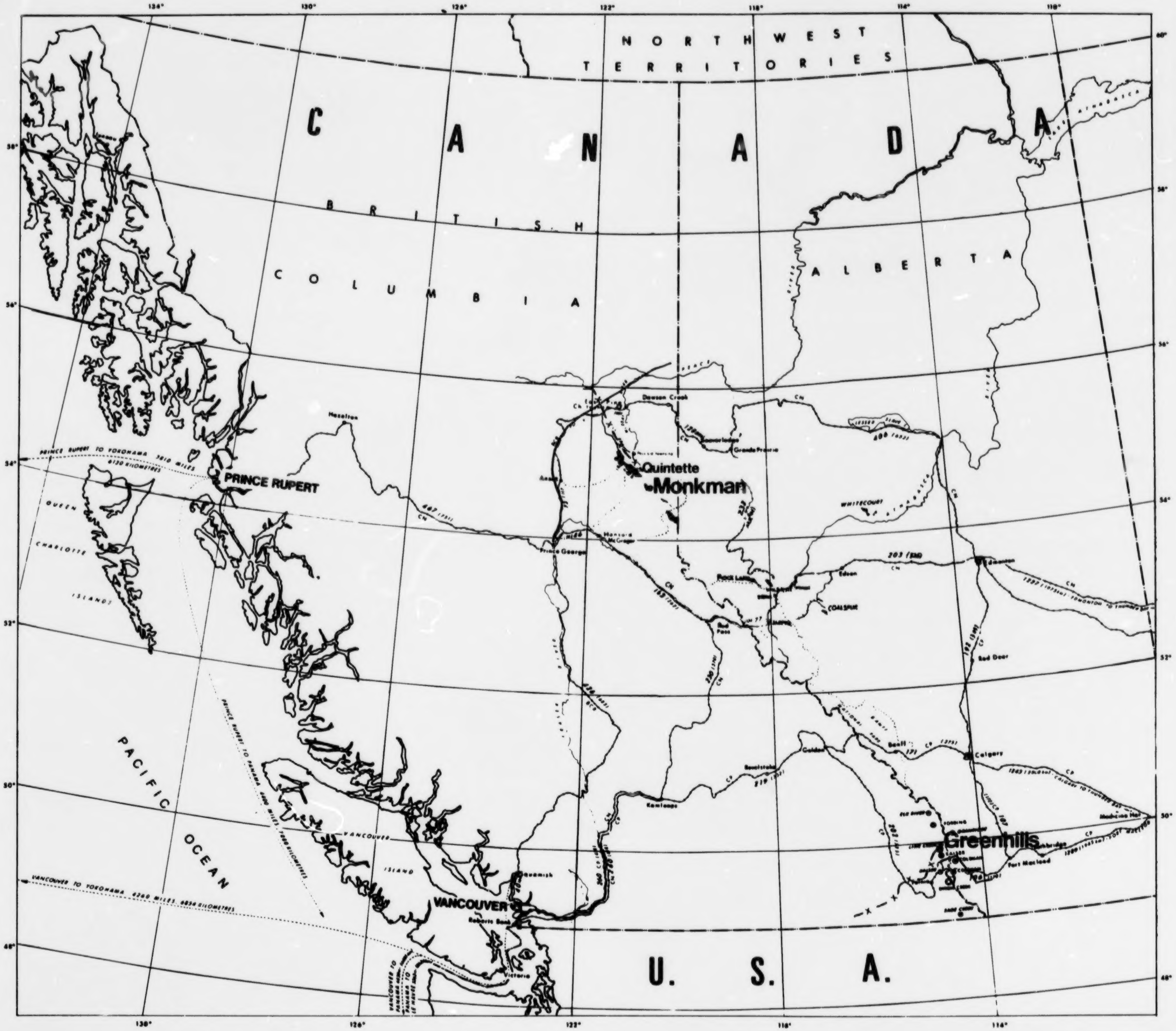
3.1.3 Analytical and Petrographic Data

The important analytical and petrographic properties relating to the liquefaction potential of the Greenhills and Monkman coals are presented in Table 3.3 and compared to Illinois No. 6 seam coal from the Burning Star Mine. Illinois No. 6 is typically used as a reference coal in liquefaction studies.

Reactive maceral content, mean vitrinite reflectance and volatile matter content are all good indications of the liquefaction potential of a particular coal. High vitrinite or reactive maceral content and low mean vitrinite reflectance indicate that the specific coal has good liquefaction potential, whereas volatile matter content indicates the fraction of the coal that can be volatilized easily.

It should be noted that the liquefaction related properties mentioned above are generalities referring to a wide range of coals. Exceptions to these generalizations will occur and in particular vitrinite reflectance is not a good measure of liquefaction potential for coals of lower rank than high volatile "B". Illinois No. 6 coal varies from the upper range of high volatile "C" (close to high volatile "B") to the lower end of high volatile "C" bituminous coal (close to sub-bituminous).

Comparison of the data in Table 3.3 enables a qualitative assessment to be made of the liquefaction potential of the nominated coals. The reactive maceral content, mean vitrinite reflectance values and volatile matter content are all consistent in the indication that Greenhills coal should offer better liquefaction potential than Monkman coal. However, both these coals would be expected to offer much less potential for liquefaction than Illinois No. 6. A quantitative assessment of the liquefaction potential of these coals is included in Section 5.0, Coal Testing Program.


NORTHWEST TERRITORIES
CANADA
BRITISH COLUMBIA
ALBERTA
U.S.A.
PACIFIC OCEAN
QUEEN CHARLOTTE ISLANDS
VANCOUVER ISLAND
PRINCE RUPERT
VANCOUVER
Victoria
Roberts Bank
Squamish
Kamloops
Revelstoke
Golden
Banff
Calgary
Medicine Hat
Lethbridge
Fort Macleod
Red Deer
Edmonton
Edson
COALSPUR
WHITECOURT
Rock Lake
Red Pass
Prince George
McBride
Dawson Creek
Grande Prairie
Hazelton
Quintette
Monkman
Greenhills
PRINCE RUPERT TO YOKOHAMA 3810 MILES 6130 KILOMETRES
VANCOUVER TO YOKOHAMA 4260 MILES 6854 KILOMETRES
PRINCE RUPERT TO PANAMA
PEACE
ATHABASCA
134°
130°
126°
122°
118°
114°
110°
60°
58°
56°
54°
52°
50°
48°

Legend

Rail-Miles (Kilometres) Between Points
Steamship Route With Distances
Existing Railroad
Possible Railroad
International Boundary
Provincial Boundary
20 0 20 40 60 80 100 Miles
20 0 20 40 60 80 100 Kilometres
Gas Pipeline
Powerline

Western Canada Coal Liquefaction Study Prefeasibility Study

Location of Potential B.C. Coal Reserves Figure 3.1

3349 21591

TABLE 3.1

GREENHILLS COAL ANALYSIS

Source: B.C. Coal Analysis of Sept. 8, 1981

Proximate Analysis, Raw Coal

Air Dry Moisture %	4.5
Ash % D.B.	13.6
Volatile Matter % D.B.	23.2
Fixed Carbon % D.B.	63.2
Sulphur %	0.36
Heating Value BTU/lb	14 140
Free Swelling Index	6.5
Plasticity	131.6 D.D.M.

Size Analysis

Size Fraction	wt. %	Cum wt%
+ 1/4"	17.3	17.3
- 1/4" x 28 m	46.0	63.3
- 28 x 35 m	7.7	71.0
- 35 x 48 m	6.5	77.5
- 48 x 60 m	2.3	79.8
- 60 x 100 m	6.2	86.0
- 100 x 0	14.0	100.0

Ultimate Analysis, Dry Basis

Carbon %	74.32
Hydrogen %	4.16
Nitrogen %	0.99
Sulphur %	0.38
Ash %	13.67
Oxygen % (by difference)	6.17
Inert Matter %	0.31

Maceral Composition Analysis

Vitrinoids (Colinite and Telinite) %	55.9
Reactive Semi-Fusinoids %	15.1
Exinoids %	0.3
Total Reactive Components	71.3

TABLE 3.1 (Cont'd)

Maceral Composition Analysis	(Cont'd)	
Inert Semi-Fusinoids %	15.1	
Macrinoids and Micrinoid %	4.0	
Fusinoids and Sclerotinite %	1.8	
Mineral Matter %	7.8	
Total Inert Components %	28.7	
Mean Reflectance Index	1.12	
Strength Index	4.26	
Compositional Balance Index	1.07	
Predicted Stability	58.3	
Hardgrove Grindability Index	107	
Ash Fusion Temperatures (°C)		
	Oxidizing	Reducing
Initial Deformation Temp.	1450	1330
Softening Temp.	1490	1425
Hemispherical Temp.	1500	1455
Fluid Temp.	1515	1480
Gieseler Fluidity		
Start	403°C	
Fusion Temp.	428°C	
Maximum Fluid Temp.	457°C	
Final Fluid Temp.	480°C	
Solidification Temp.	487°C	
Melting Range	52°C	
Maximum Fluidity	131.6 dd/m	
Ash Mineral Analysis		
SiO_2	59.78%	
Al_2O_3	28.03%	
TiO_2	1.62%	
Fe_2O_3	4.15%	
CaO	1.68%	
MgO	0.88%	
Na_2O	0.35%	
K_2O	0.90%	
P_2O_5	0.97%	
SO_3	0.62%	
Undetermined	1.02%	

TABLE 3.2

MONKMAN COAL ANALYSIS

Source: Birtley Coal and Minerals Testing, Lab. No. 8883

Proximate Analysis, Raw Coal

Air Dry Moisture %	4.9
Ash % D.B.	19.3
Volatile Matter % D.B.	20.4
Fixed Carbon % D.B.	59.5
Sulphur %	0.4
Heating Value BTU/lb	14 400
Free Swelling Index	4

Size Analysis

Size Fraction	wt. %	Cum wt%
+ 1/4"	28.7	28.7
- 1/4" x 28 m	47.4	76.1
- 28 x 48 m	7.5	83.6
- 48 x 100 m	6.3	89.9
- 100 x 0	10.1	100.0

Ultimate Analysis

Moisture	0.78
Carbon %	72.68
Hydrogen %	3.86
Nitrogen %	0.73
Sulphur %	0.40
Ash %	19.3
Oxygen % (by Difference)	2.25

Maceral Data

Vitrinoids %	57.08
Exinoids %	-
Semifusinite %	31.81
Macrinite %	-
Micrinite %	1.74
Fusinite	1.96
Mineral Matter %	7.41
Total Reactives %	65.37
Inerts %	34.63

TABLE 3.2 (Cont'd)

Maceral Data (Cont'd)

Mean Reflectance Index	1.2784
Strength Index	6.34
Compositional Balance Index	1.81
Predicted Stability	62
Hardgrove Grindability Index	81

Ash Fusion Temperatures (°C)

	Oxidizing	Reducing
Initial Deformation Temp.	1330	1250
Softening Temp.	1480	1410
Hemispherical Temp.	1495	1440
Fluid Temp.	1505	1475

Gieseler Fluidity

	DDPM	Temp. °C
Start	1	447
Maximum	4	469
Final	0	502
	Range	55

TABLE 3.3

COMPARISON OF GREENHILLS, MONKMAN AND ILLINOIS NO. 6 COALS

Source: Greenhills and Monkman as per Tables 3.1 and 3.2
Illinois No. 6, courtesy of HRI

	Greenhills	Monkman	Burning Star Illinois No. 6 Typical
Proximate Analysis, Raw Coal			
Moisture %	4.5	4.9	10
Ash % D.B.	13.6	19.3	11.4
Volatile Matter % D.B.	23.2	20.4	37.5
Fixed Carbon % D.B.	63.2	59.5	51.0
Calorific Value (Dry), kCal/kg	32 800	33 400	28 300
Ultimate Analysis, Dry Basis			
Carbon %	74.3	73.2	68.4
Hydrogen %	4.2	3.9	4.3
Nitrogen %	1.0	0.7	1.4
Sulphur %	0.4	0.4	3.4
Ash %	13.7	19.5	11.4
Oxygen %	6.2	2.3	11.1
Others %	0.3	-	-
Petrographic			
Reactive Macerals %	71	65	84
Mean Vit Reflectance	1.12	1.28	0.6 - 0.7

4.0 DESCRIPTION OF PROCESSES

4.1 GENERAL

The primary objective of any coal liquefaction process is to increase the hydrogen to carbon ratio from that of the coal feed (typically $CH_{0.4}$ - $CH_{0.6}$) to that of premium liquids (typically CH_2). However, although all liquefaction processes produce somewhat similar products, the manner in which these products are manufactured differs considerably, depending on the type of process used.

The four principal types of coal liquefaction process are pyrolysis, direct liquefaction, indirect liquefaction and solvent extraction. Of the five processes selected by the Western Canada Coal Liquefaction Study Consortium for comparison in this study, four of these may be classed as direct liquefaction processes and one as a solvent extraction process viz:

H-Coal	Direct Liquefaction
Ruhrkohle/Veba	
SRC-II	
EDS	
NCB/BP	Solvent Extraction/Hydrocracking

Direct Liquefaction

In direct liquefaction processes the coal is finely ground then mixed with a liquid recycled from the process. The slurry is pressurized to between 14 000 and 21 000 kPa and hydrogen is added. The mixture of gas, liquid and solid is preheated to between 300 and 400°C at which temperature the coal particles start to dissolve in the solvent and react with the hydrogen. The hydrocarbon products formed vary in range from light gases through to heavy liquids, and these must then be separated from the remaining solids, principally unreacted coal and mineral matter. The solids-free liquids are recovered by flashing the slurry to a low pressure followed by conventional atmospheric and vacuum distillation.

The product slates from direct liquefaction type processes are all somewhat flexible in the relative proportions of liquid products. Within limits specific products can be maximized by varying operating conditions, but each process would still produce SNG, LPG, naphtha, middle distillate and heavy distillate. The SNG and LPG are saleable products, but the naphtha and heavier fractions require substantial upgrading prior to use as gasoline, petrochemical naphtha, jet fuel or diesel fuel. Upgrading technology to meet these product specifications is currently underway.

Direct liquefaction processes are further classified as catalytic or non-catalytic. H-Coal and Ruhrkohle/Veba are catalytic processes since the hydrogenation of the coal occurs in the presence of a catalyst. In the H-Coal process the semi-precious metal catalyst is kept in a fluidized state, known as an ebullated bed. In the Ruhrkohle/Veba process the catalyst, known as Red Mud, is dispensible. The presence of a catalyst should assist hydrogenation of the coal, so reducing the sensitivity of the process to variations in coal quality.

SRC-II and EDS are classified as non-catalytic processes since coal hydrogenation proceeds without the presence of a catalyst. SRC-II relies on the catalytic activity of the coal ash to promote the hydrogenation. This makes the process very sensitive to coal quality since different coals contain ashes with different catalytic properties. In the EDS process the process derived solvent is catalytically hydrogenated prior to mixing with the coal. This enhances the hydrogenation of the coal, thus permitting the EDS process to be less sensitive to coal quality than SRC-II.

Solvent Extraction/Hydrocracking

In the NCB/BP Liquid Solvent Extraction (LSE) process, liquefaction of the coal occurs at much lower pressures than in the direct processes. However, the first step of the process is similar to that of all the direct liquefaction processes in that the finely ground coal is slurried with a liquid recycled from the process. The slurry is pressurized to between 1 400-2 100 kPa, heated to about 400°C then fed to a digester. In the digester the residence time is sufficient to permit the organic portion of the coal to be almost completely extracted into solution. The undissolved material, mainly minerals, ash and unreactive coal, is separated from the extract by filtration.

Once the solids have been removed, the extract can be hydrocracked using a catalyst specific to the desired product slate. The hyrocracking operation is usually similar to that of conventional petroleum refining occuring at 14 000-21 000 kPa. This yields a product slate similar to that from the direct liquefaction processes namely SNG, LPG, naphtha and middle distillate. Depending on the type of coal feed, heavier fractions may also be obtained which may be consumed within the complex.

Although the NCB report that 90-95% of the coal macerals can be extracted, the extract yield may vary considerably depending on coal quality. The liquid yields are also enhanced by the mildly hydrogen donating properties of the solvent recycled from the hydrocrackers.

4.2 H-COAL PROCESS

4.2.1 Introduction

H-Coal is a direct catalytic hydroliquefaction process developed by Hydrocarbon Research Inc. to convert coal into a variety of synthetic liquid fuels ranging from distillate syncrude to heavy fuel oil. The process uses the commercially proven ebullating bed reactor and a Co-Mo catalyst to achieve 40-50 weight percent C_4-525°C distillate yields (on a dry basis) from a wide variety of coals.

4.2.2 Process Description

In the H-Coal process the coal is washed, crushed to 30 mesh, or lower size, dried and slurried with a process derived recycle oil, pumped to a reactor pressure of 21 000 kPa, mixed with hydrogen, preheated to 400°C and fed to the reactor. The reactor operates at approximately 450°C and 21 000 kPa and extra heat required to bring the feed up to its temperature is provided by the reaction (an advantage over comparable direct liquefaction processes operating under similar conditions which have experienced preheater coking problems due to higher temperatures in the preheater). The feed and recycle from the ebullating pump produces a fluidized catalyst bed in the reactor. Catalyst addition and removal on a continuous basis is performed in order to maintain its high activity (2-3% catalyst on a daily basis amounting to a consumption of approximately 0.5 kg catalyst/tonne of dry coal). Reactor products, including ash and unconverted coal, are separated from the gases by flashing in a series of vapour-liquid separators.

C_1-C_4 are recovered from the gaseous products. Hydrogen recovery from the vapour products is maximized at the intermediate and low pressure levels. The reactor slurry is processed through hydroclones to reduce its solids content. Low solid content overflow is recycled as a slurry oil for the feed coal while the underflow is fractionated to produce distillate products and residue.

Depending on the process severity selected, the product yield can be all distillate material in the "Syncrude" mode (high severity) or a distillate and a heavy fuel oil in the "Fuel Oil" mode (low severity). Typical hydrogen consumption is 5.5 kg and 3.5 kg per 100 kg dry coal for the Syncrude and Fuel Oil modes, respectively.

A typical product slate with Illinois No. 6 coal in the Syncrude mode consists of 12-14 percent C_1-C_4, 18-20 percent C_5+ naphtha fraction, 28-32 percent middle plus heavy distillate fraction and 32-34 percent residue, on an maf coal basis.

In the Fuel Oil mode, the process concept is that ash, unconverted coal and preasphaltenes will be separated from the product in the hydroclones underflow by a heavy naphtha antisolvent. This step is reported to be under development at the present time.

The distillate products are stabilized by hydrotreating at different severity levels. Details on the stage of development of raw product upgrading options are given in Section 4.7.

The vacuum residue is fed to a partial oxidation unit using a Texaco slagging gasifier to satisfy the complex hydrogen requirements. Information on the successful demonstration of the Texaco gasifier on H-Coal residue is given in Section 4.7.

4.2.3 Process Development

H-Coal has been under development for more than fifteen years evaluating nineteen types of coal. Development and demonstration have been carried out on a 12 kg/day bench scale unit and in a Process Development Unit (PDU) handling 3.0 tonne of coal per day. The bench scale units are used for process improvement studies, catalyst evaluation and testing of new coals, while the PDU studies aim at confirming the design basis, operating conditions and modes of operation.

Coal preparation and hydrogenation, vapour-liquid separation, solid-liquid separation, hydrogen recovery, oil recycle and fractionation have been tested in the PDU.

The process feasibility is being currently demonstrated at the 180 (540) tonne/day pilot plant in Catlettsburg, Ky., as the last step to commercialization. Performance and product yields will be confirmed and commercial size equipment will be proven during the plant operations.

The units being tested at Catlettsburg are coal receiving, coal preparation and hydrogenation, vapour-liquid separation, liquid product recovery, hydrogen recovery, oil recycling and sour water treating. Bottom residue processing, sour gas treatment and hydrogen manufacture are not included.

The HRI development program is presently sponsored by the U.S. Department of Energy, EPRI, Ashland Oil Inc. (the operator of the 180 tonne/day pilot plant), Standard Oil of Indiana, Conoco Coal Development, Mobil Oil, the Commonwealth of Kentucky and Ruhrkohle AG.

The bench scale unit has successfully processed the following coals:

Bituminous:
Black Mesa, Colorado, German "Steinkohle", Illinois #6, Kentucky #9, Kentucky #11, Kentucky #14, Pittsburgh seam, Annesley Double Coal (U.K.), Utah D, Daiparowits Plateau.

Sub Bituminous:
Wyodak, Big Horn.

Lignite:
North Dakota, Texas Lignite.

Brown:
Australian Brown.

Ten runs of approximately 30 days each have been completed with Illinois #6, Wyodak and Kentucky #11 in the PDU.

Continuous catalyst addition and removal, syncrude and fuel oil operations and a two stage letdown system have been demonstrated. Process fluid dynamics for reactor scale up and critical operating parameters have been established.

The pilot plant commenced a two year demonstration program in May 1980. Tests in the Syncrude mode have been carried out to date on Illinois #6, and Kentucky #11 coals, their main goals being to prove the mechanical operability of commercial equipment, determine its maintenance requirements, test materials of construction, verify yields, provide products for commercial testing and check scale up and engineering data.

A 45 day syncrude operation on Illinois #6 was completed during February-March 1981. It was interrupted by pipe blockages originating from an operational problem. A run on Kentucky #9 coal started in May 1981 was stopped due to a failure of a charge pump which caused coking in the reactor. The run was successfully completed in June and July 1981.

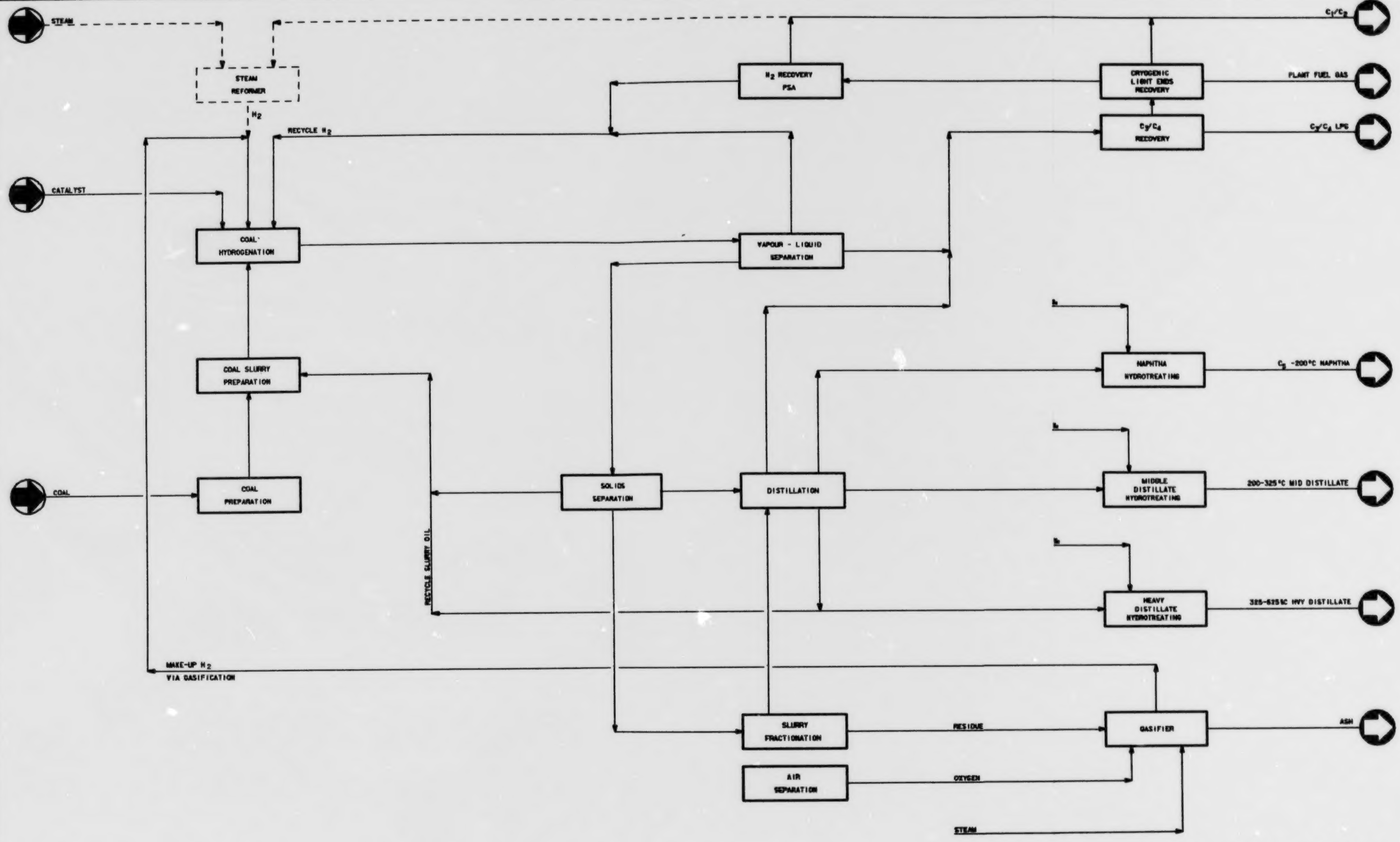
A highly successful run of 136 days was completed in December of 1981 on Illinois #6 coal in the Syncrude mode. The run was terminated as all major program objectives had been achieved, not because of mechanical problems. During the 136 day run, the pilot plant was on coal approximately 75% of the time.

Another problem encountered in the pilot plant was erosion in the high pressure letdown valves. The trim erosion problem has been solved by proper choice of materials and the packing materials are being improved at present.

4.2.4 Commercialization Plans

The U.S. Department of Energy has authorized the beginning of design work on a commercial scale H-Coal plant to be located at Breckenridge, Kentucky. The 20 000 tonne/day run of mine plant based on Illinois #6 coal feed will produce 50 000 BBL/day of syncrude fuels and 30 MM SCF/day of SNG. Main products from the Breckenridge Project will include propane, butane, gasoline blend stock (C_5 - 180°C) and distillate fuel oil (180 - 510°C). It is understood that the distillate fuel oil has been hydrotreated sufficiently to meet industrial turbine fuel and possibly home heating oil specifications. Phase zero including design, cost estimate and economic evaluation and detailed plans for construction and operation has just been completed by Bechtel (funded by DOE, Ashland Oil and AIRCO).

Detailed engineering is tentatively scheduled for the first quarter of 1982, construction commencing in 1984 and start-up scheduled for mid 1986. This is subject to financial support by the Synthetic Fuels Corporation. The plant configuration includes eight reactor trains (one of them spare) and two downstream fractionation units.


STEAM
STEAM REFORMER
H_2
RECYCLE H_2
CATALYST
COAL HYDROGENATION
COAL SLURRY PREPARATION
COAL
COAL PREPARATION
RECYCLE SLURRY OIL
MAKE-UP H_2 VIA GASIFICATION
H_2 RECOVERY PSA
VAPOUR - LIQUID SEPARATION
SOLIDS SEPARATION
DISTILLATION
SLURRY FRACTIONATION
AIR SEPARATION
RESIDUE
OXYGEN
STEAM
GASIFIER
ASH
CRYOGENIC LIGHT ENDS RECOVERY
C_3/C_4 RECOVERY
C_1/C_2
PLANT FUEL GAS
C_3/C_4 LPG
H_2
NAPHTHA HYDROTREATING
C_5 -200°C NAPHTHA
H_2
MIDDLE DISTILLATE HYDROTREATING
200-325°C MID DISTILLATE
H_2
HEAVY DISTILLATE HYDROTREATING
325-525°C HVY DISTILLATE

Western Canada Coal Liquefaction Study
Prefeasibility Study
Simplified Block Flow Diagram
H-Coal Process

4.3 RUHRKOHLE/VEBA PROCESS

4.3.1 Introduction

The production of distillate oil from coal based on a catalytic process was developed in Germany by Bergius and Pier in the early twentieth century. This process, better known as the I.G. Farben technology, was used extensively to fulfill Germany's requirements of gasoline and middle distillates during World War II.

After a lapse of several years, Ruhrkohle AG and Veba Oel AG jointly set out to modify the I.G. Farben technology (German Process), by introducing process steps and equipment to the most recent state of technology, while still utilizing the same cheap and abundantly available iron oxide (Red Mud) catalyst to promote the liquefaction reactions. The modifications were made to develop industrial scale capability for the production of liquid products from coal with improved economic conditions relative to the technology developed in Germany during the war.

The development has reached the stage of pilot plants, which are designed by extrapolation of laboratory plants and bench scale units. It is anticipated that the pilot plant data would be adequate to design commercial scale plants with minimum risks.

4.3.2 Modifications of the German Process

The three main modifications made to the I.G. Farben process are the use of the blend of middle oil and heavy oil from the process as a solvent, the separation of solids and asphaltenes by distillation and the production of hydrogen from the residue containing solid materials and asphaltenes. These modifications have facilitated the reaction to be carried out at a substantially reduced pressure of 30 000 kPa instead of the 70 000 kPa reactor pressure previously used. They also increased the specific coal throughput by 50%, increased the thermal efficiency, improved the heat recovery and reduced the conversion costs. The improved yields and operating conditions have been confirmed in a process development unit.

4.3.3 Process Description

In the Ruhrkohle/Veba process, the dry washed coal is ground to <0.2 mm, mixed with 1-2 wt% iron oxide (Red Mud) catalyst and slurried in a 40% middle distillate and 60% heavy oil recycle stream to give a 40-50% coal-in-oil slurry. The coal-oil slurry is pressurized to 32 000 kPa, mixed with make-up and recycle gas (hydrogen) and preheated via feed/product heat exchange trains to about 380°C. The slurry is further heated up to 425°C by fired heaters before introduction to the hydrogenation reactors.

The hydrogenation reactor trains consists of three upflow reactors in series, operating at 30 000 kPa pressure and 475°C temperature with a CHSV of approximately 0.5 tonne maf coal/m^3 reactor volume per hour. Quench hydrogen is fed into each reactor to control the exothermic reaction and prevent the reaction temperature from exceeding 475°C. Each reactor is about 2.4 m I.D. and 18 m high with an internal refractory lining. There are no other internals to interfere with backmixing which is created by the upward hydrogen flow through the slurry.

The reactor products are separated in a series of flash separators. Reactor gases and vapours are removed from the solid-liquid phase in a hot, high pressure separator, cooled in the feed/product heat exchange train and a direct water quench (to control ammonia salts deposition) and phase separated to give a recycle hydrogen stream, a coal oil stream and a waste water stream. The recycle hydrogen stream is scrubbed in an oil scrubber to recover part of the light hydrocarbons (C_1-C_5) and enrich the hydrogen stream for recycle. The excess scrubbed gas is discharged from the system as purge gas.

The coal oil fraction from the separation section is processed in an atmospheric distillation unit for the recovery of raw naphtha (C_5-200°C), middle distillate (200-325°C) and heavy distillate (>325°C) fractions. A gasoline stabilizer and light ends recovery system are provided for C_3/C_4 LPG recovery.

The solid-liquid phase from the hot separator is expanded through two or three stages of let-down valves and flash vessels to remove remaining gases and vapour, then vacuum distilled into heavy vacuum oil and residue fractions. The heavy vacuum oil, most or all of the heavy distillate oil from the atmospheric distillation unit and a portion of the middle distillate are recycled as slurry oil.

The residue, containing minerals, undissolved coal, catalyst and heavy hydrocarbons (mainly >540°C asphaltenes, preasphaltenes and resins), is gasified via partial oxidation to produce part of the hydrogen for the hydrogenation reaction and subsequent upgrading of the products.

The raw products require hydrotreating to produce stable products that can be blended to produce a high quality coal derived synthetic crude or marketed individually. Bench scale hydrotreating and hydrocracking test programs, conducted by Veba Oel AG and BASF Ludwigshafen, have demonstrated that the raw products produced by the German process can be upgraded successfully to finished products. Hydrotreating of the raw naphtha fraction produces an acceptable catalytic reformer feedstock which on subsequent reforming produces gasoline with RON of 100-103. Hydrotreating of the middle distillate fraction produces an industrial grade fuel oil. Mild hydrocracking of the hydrotreated middle distillate fraction produces acceptable quality diesel or jet fuel. Hydrotreating of the heavy distillate fraction (if any net product is produced) produces a low sulphur industrial fuel oil or an acceptable catalytic cracker feedstock.

4.3.4 Process Development

Extensive process development has been carried out by Ruhrkohle/Veba since 1975 on a 250 kg/day maf coal process development unit operated by Bergbau-Forschung Gmbh. Many coals have been tested on the PDU varying in rank from lignite to bituminous. The results of the testing were sufficiently successful for Ruhrkohle AG and Veba Oel AG to design and construct a 200 tonne/day maf coal pilot plant at Bottrop, F.R.G. Commissioning of this pilot plant is currently underway. The current sponsors of the pilot plant program are the government of the Federal Republic of Germany, Ruhrkohle AG and Veba Oel AG with Ruhrkohle Oel und Gas Gmbh as the operator.

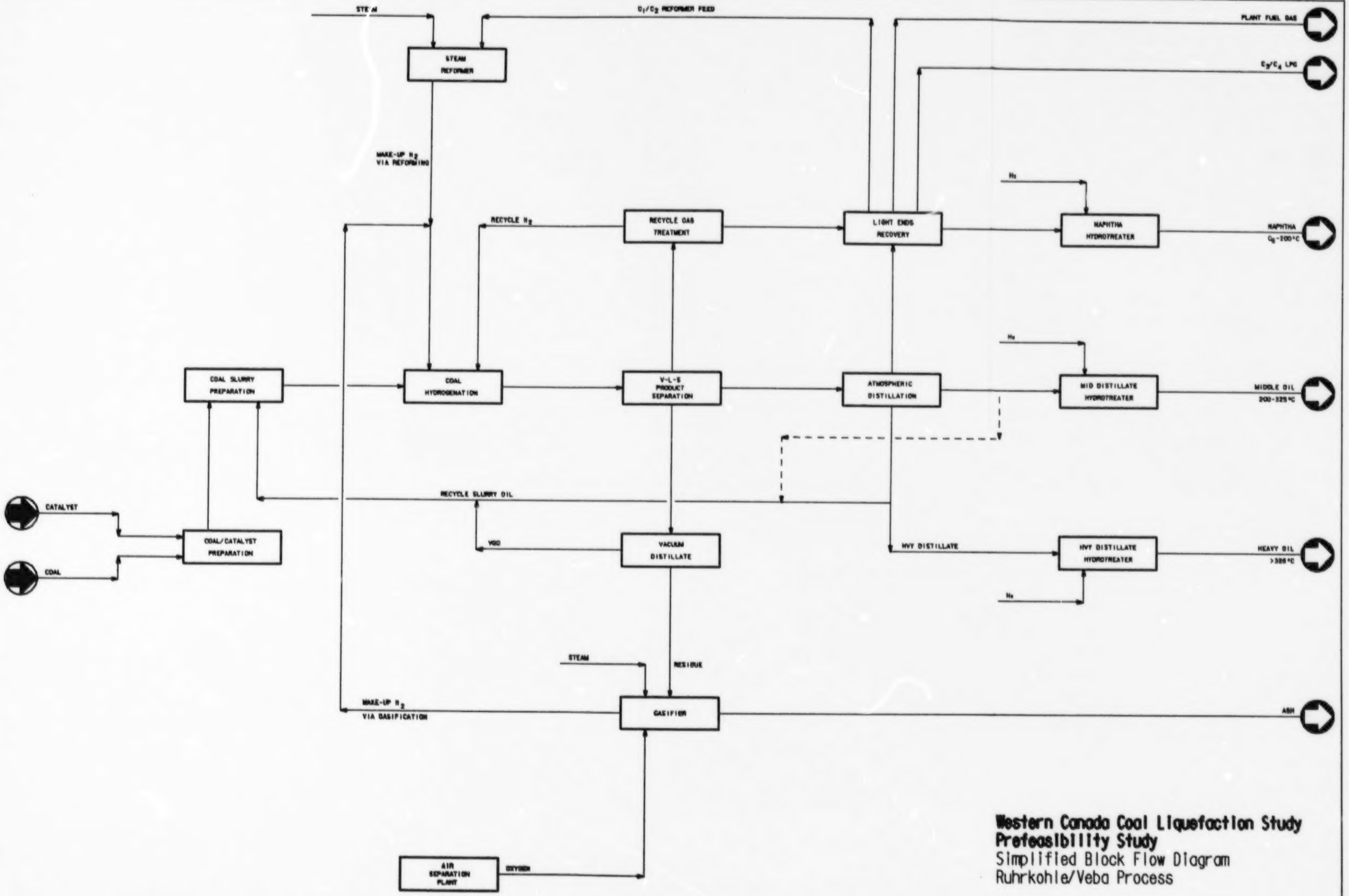
The pilot plant includes as process units: coal and catalyst handling, hydrogenation, atmospheric distillation, gas treating, vacuum distillation, high pressure residue flashing and residue granulation. A three year test program for the pilot plant is reported to include about 4 000 hours of testing in 1981, 5 000 hours in 1982 and 6 000 hours in 1983. It is evident, however, that this schedule will have to be modified as plant commissioning is now underway and "first coal in" occured only in January 1982. Ruhrkohle/Veba report that coal processing should commence in the first quarter of 1982. Critical design areas to be investigated are coking in the feed heat exchange preheat train and feed fired heater, reactor refractory design and hydrogen containment and the high pressure solid-liquid letdown system.

An in-depth test program to study upgrading alternatives for naphtha, middle distillate and heavy oil is underway and will proceed in conjunction with the pilot plant tests. Preliminary testing has shown that maximum utilization of the naphtha would include hydrotreating and reforming to yield gasoline, further aromatic extraction to yield BTX and finally steam-cracking for olefin production. Middle distillate upgrading options include hydrotreating for fuel oil, followed by either low severity hydrocracking for diesel and jet fuel or high severity hydrocracking for naphtha. Heavy distillate upgrading options include hydrotreating for industrial fuel oil and high severity hydrocracking for diesel, jet fuel and naphtha. Additional details are given in Section 4.7 on the stage of development of upgrading options via hydrotreating and hydrocracking.

A residue gasification test program is also planned at the nearby Ruhrchemie/Texaco partial oxidation pilot plant to confirm residue gasification technology.

4.3.5 Commercialization Plans

Four commercial scale coal liquefaction facilities are currently being proposed in Germany by Ruhrkohle AG, Veba Oel AG, Saarbergwerke AG and Rheinische Braunkohlenwerke AG, all utilizing versions of the IG Farben technology. The proposed plants, capable of processing between 6 000 and 20 000 tonne (maf) coal per day, will produce motor fuels, chemical raw materials and light fuel oils. Technical designs, costs, locations and permits, feeds and products and economics are being investigated by each company for possible construction and startup by 1990. As each project is dependent on financial support by the German Government, it is likely that not all projects will go foward. However, one plant may be built for security of oil supply reasons.


STEAM
C_1/C_2 REFORMER FEED
PLANT FUEL GAS
STEAM REFORMER
C_3/C_4 LPG
MAKE-UP H_2 VIA REFORMING
H_2
RECYCLE H_2
RECYCLE GAS TREATMENT
LIGHT ENDS RECOVERY
NAPHTHA HYDROTREATER
NAPHTHA
C_5-200°C
H_2
COAL SLURRY PREPARATION
COAL HYDROGENATION
V-L-S PRODUCT SEPARATION
ATMOSPHERIC DISTILLATION
MID DISTILLATE HYDROTREATER
MIDDLE OIL
200-325°C
RECYCLE SLURRY OIL
CATALYST
COAL/CATALYST PREPARATION
COAL
VGO
VACUUM DISTILLATE
HVY DISTILLATE
HVY DISTILLATE HYDROTREATER
HEAVY OIL
>325°C
H_2
STEAM
RESIDUE
MAKE-UP H_2 VIA GASIFICATION
GASIFIER
ASH
AIR SEPARATION PLANT
OXYGEN

Western Canada Coal Liquefaction Study
Prefeasibility Study
Simplified Block Flow Diagram
Ruhrkohle/Veba Process

4.4 NCB/BP LIQUID SOLVENT EXTRACTION PROCESS

4.4.1 Introduction

The British National Coal Board have been conducting research into coal extraction for approximately thirty years. More recent process development has been directed towards production of premium transportation fuels by liquid or super critical gaseous extraction of coals followed by hydrocracking of the extract. Of these two processes the liquid solvent extraction (LSE) process is economically more attractive and also gives better yields.

The LSE process is a two stage process, unique among the other direct liquefaction processes, in that production of the coal extract is separated from the subsequent hydrogenation and cracking of the extract. Approximately ten years of extraction testing has enabled the NCB to optimize conditions such that they now claim up to 95% of the maceral content of coal can be extracted. This extraction rate varies with the type of coal feed. NCB claim that small scale testing has demonstrated that the extract can be successfully separated from the undissolved material and subsequently hydrocracked.

NCB/BP claim this sequential operation is particularly suited to the production of transportation fuels since only the filtered or "clean" coal extract (free from unreacted coal, ash and mineral matter) is treated in the hydrocracking stage. Thus, since the feed to the hydrocracker has had many potential "poisons" removed, the catalyst can be selected to optimize the production of premium distillate products. This scheme also restricts hydrogenation to the selected lighter components of the coal, presumably minimizing hydrogen consumption.

4.4.2 Process Description

Although there is no maximum limit on the coal ash content, beneficiation in a coal preparation plant may prove economically advantageous. After any necessary beneficiation process, feed coal is finely crushed and dried prior to mixing with a heavy oil solvent, recycled from a downstream fractionation facility.

The slurry of recycle oil and crushed coal is pressurized to about 2 000 kPa and rapidly heated to above 400°C before entering the extraction vessels. These are sized so that the residence time (typically between 30-60 minutes) permits complete extraction of the coal maceral content. Hydrogenation and solvation of the macerals is accomplished using the properties of the recycle oil solvent. This heavy oil solvent is a portion of the hydrocracker liquid product stream and so acts as a mild hydrogen donor.

The coal digest from the extractors is cooled then flashed to about 300 kPa for gas/slurry separation. The gas stream is cooled to condense light liquids then sent to the gas processing section. The slurry stream contains ash and coal residues which are separated from the hydrocarbon liquids by filtration. Fast filtration is stated by NCB to be possible since extraction is carried out under optimized conditions. The solid residue, after washing to recover any entrained extract, may then be fed to the Gasifier to produce hydrogen and/or used as a boiler fuel.

The extract solution is pressurized to about 20 000 kPa then mixed with hydrogen-rich gas. This combined stream is heated to approximately 420°C in preparation for hydrocracking. The hot hydrocracker feed may require additional purification by passage through one or two guard reactors prior to entering the hydrocracker reactor. The hydrocracking reactors have fixed beds of catalyst with provision for the addition of quench gas between the beds. Hydrocracker product is cooled by feed/ product interchange and by steam generation.

Cooled product at 300°C is separated into gases and liquids by multi-stage flashing and fractionation. The gases liberated are sent to the gas processing section while the liquid stream is stabilized in a debutanizing column. The stabilized liquid is then fractionated in atmospheric and vacuum columns. This fractionation is designed to produce an overhead naphtha cut (C_5-175°C), a middle distillate cut (175-300°C), the heavy oil recycle solvent and the residual high boiling oils which are used as fuel oil. The naphtha and middle distillate require conventional petroleum type hydrotreating and fractionation to reduce the nitrogen, sulphur and oxygen contents. This naphtha may be subsequently reformed to yield a high quality gasoline while it is reported that the hydrotreated mid-distillate can yield jet and diesel fuels.

The hydrogen requirements for hydrocracking and hydrotreatment would normally be produced by oxygen blown gasification of the solid residue from the filtration step. Alternatively, part or all of the residue could be used as boiler fuel in the steam and power generation plant. In this case hydrogen would be provided by steam reforming of imported natural gas or by gasification of additional coal.

Hydrocracker and digestor offgases and debutanizer overhead product are treated in a gas processing section for recovery of acid gases, hydrogen, LPG and process fuel. Acid gases are removed via conventional amine scrubbing and the H_2S is converted to elemental sulphur. Hydrogen is cryogenically separated from the gas stream, compressed and combined with make-up and hydrocracker recycle hydrogen streams. LPG product is recovered by fractionation and the remaining gases, primarily C_1 and C_2, are used as process fuel.

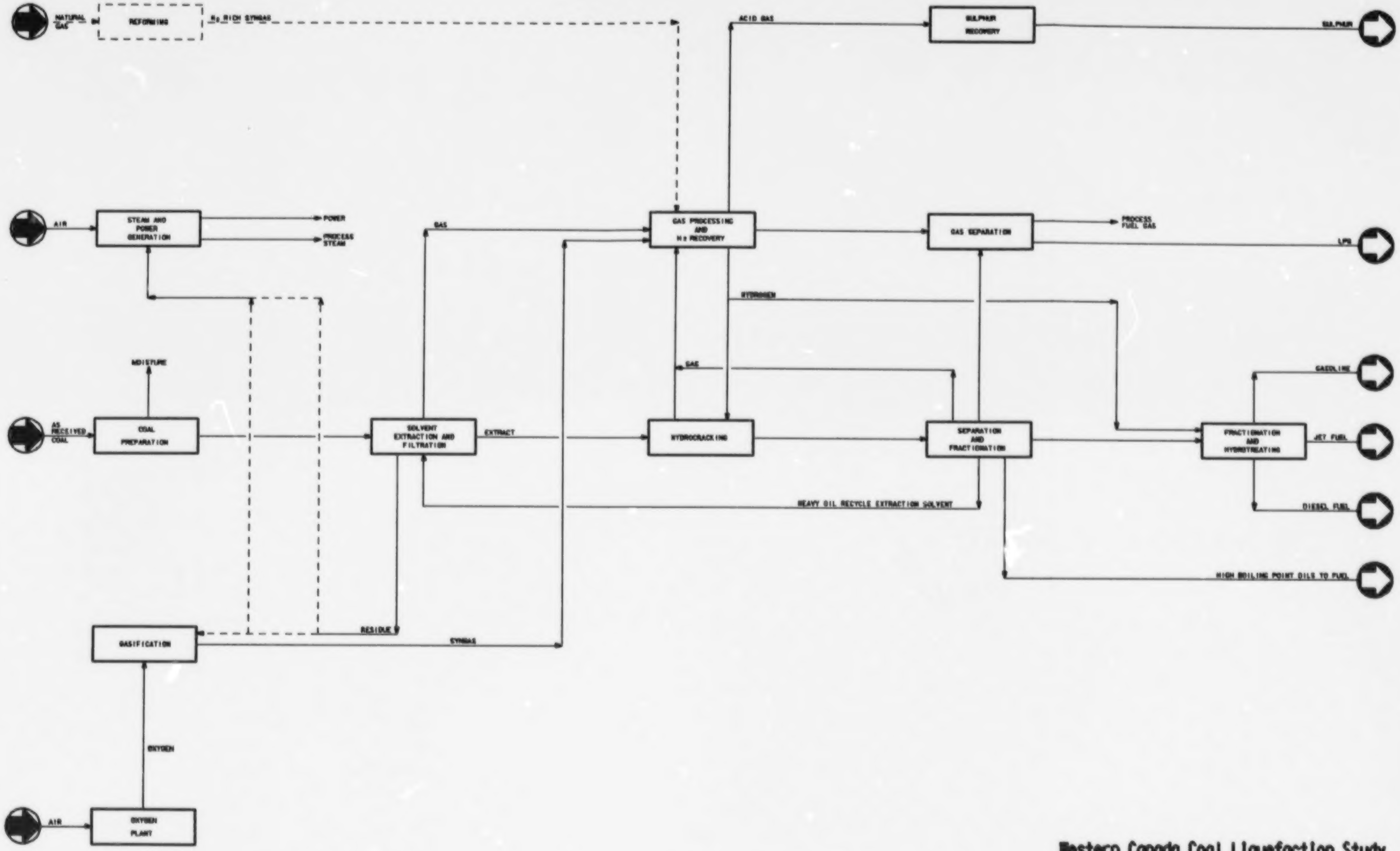
4.4.3 Process Development

A semi-continuous 30 kg/hr extraction unit has been operated by the National Coal Board for almost ten years. Many types of coal have been tested varying in rank from lignite to bituminous. Considerable testing of the coal extracts has also taken place on a bench scale hydrocracker. More than 50 different catalysts were tested by British Petroleum during the research program. These testing facilities have now been combined into an integrated unit which can process 30 kg/hr coal. The solids are removed from the extract in batch operated leaf and candle filters.

Design studies for an integrated 25 tonne/day plant have been conducted by NCB/BP and plans are proceeding for funding the detailed design and construction of this plant. Partial government funding has been obtained for the construction and operation of the 25 tonne per day pilot plant, but it is conditional on continued private industry participation (by British Petroleum and Phillips Petroleum). Detailed engineering will not commence until private industry involvement is guaranteed. Barring lengthy delays this plant could be in operation by 1984/85. After this date, NCB/BP require a three year test program to establish the design basis for a commercial plant of 5 000 tonne/day capacity.

4.4.4 Commercialization Plans

As the integrated liquid solvent extraction/hydrocracker process is in an early stage of development, no firm commercialization plans have been announced. NCB anticipate, however, that the first 5 000 tonne per day unit could be operational by the early 1990's.


NATURAL GAS
REFORMING
H_2 RICH SYNGAS
AIR
STEAM AND POWER GENERATION
POWER
PROCESS STEAM
MOISTURE
AS RECEIVED COAL
COAL PREPARATION
SOLVENT EXTRACTION AND FILTRATION
GAS
EXTRACT
RESIDUE
SYNGAS
GASIFICATION
OXYGEN
AIR
OXYGEN PLANT
GAS PROCESSING AND H_2 RECOVERY
ACID GAS
SULPHUR RECOVERY
SULPHUR
GAS SEPARATION
PROCESS FUEL GAS
LPG
HYDROGEN
GAS
HYDROCRACKING
SEPARATION AND FRACTIONATION
FRACTIONATION AND HYDROTREATING
GASOLINE
JET FUEL
DIESEL FUEL
HEAVY OIL RECYCLE EXTRACTION SOLVENT
HIGH BOILING POINT OILS TO FUEL

Western Canada Coal Liquefaction Study
Prefeasibility Study
Simplified Block Flow Diagram
NCB/BP Process

4.5 EXXON DONOR SOLVENT PROCESS

4.5.1 Introduction

Exxon Corporation conducted research into coal liquefaction technology for approximately ten years before deciding to develop the Exxon Donor Solvent (EDS) process. The unique factor of the EDS process is that the recycle solvent used to slurry the coal is hydrogenated separately in a catalytic process. This enhances the hydrogen donation properties of the solvent so facilitating hydrogenation of the coal. The decoupled configuration of the liquefaction and catalytic hydrogenation sections should also permit wide variation of product distribution by variation of conditions in either or both sections.

The original process scheme incorporated a Flexicoker to process the vacuum residue stream. This step was conceived as improving liquid yields and producing plant fuel gas. The Flexicoking operation is no longer considered the preferred method of processing the vacuum residue and investigation of different schemes is now in progress.

4.5.2 Process Description

The crushed coal is slurried with the recycle solvent then dried. The slurry is pressurized to approximately 14 000 kPa, then mixed with hydrogen. The mixture is preheated to 400°C then flows to the reactor. This is a non-catalytic, tubular upflow reactor sized to give a residence time of 20-60 minutes. The hydrogenated products are recovered as gas, liquid and solid streams by a multistage flash separation followed by atmospheric and vacuum distillation.

Naphtha is the overhead product from the Atmospheric Column and a 180°C-340°C sidestream cut is used as a portion of the recycle solvent. Atmospheric bottoms are fed to the Vacuum Column from which an upper sidestream fraction is taken as make-up for the recycle solvent. Heavy distillate is taken as the lower sidestream fraction while the vacuum bottoms residue consists of all material boiling above 510°C (ash, unconverted coal, high boiling oils, etc).

The vacuum residue may be processed in a number of different ways:

(a) Gasification to produce hydrogen and/or fuel gas. Choice of this configuration largely depends on availability of natural gas to the complex. If sufficient low price methane is available, a steam reforming unit could be preferable to a gasification scheme on account of the lower capital cost. Exxon have scheduled gasification testing with Texaco and Shell-Koppers in 1982.

(b) Flexicoking to produce primarily a low BTU fuel gas and some additional C_1/C_2 and liquid products. The vast majority of these additional liquids are high boiling coker gas oils and the low BTU fuel gas requires the use of specially designed furnaces. Consequently this configuration appears difficult to justify considering the high capital investment required.

(c) Recycling to the liquefaction section. Exxon are currently testing this mode of operation since it appears to offer greater yields of lighter products. This scheme should also offer considerable product slate flexibility but Exxon require more research and testing before quantifying the advantages.

(d) Use of the residue as a fuel in a hybrid boiler raising both process heat and steam. This scheme appears to offer favourable economics and a testing schedule is to be prepared by Combustion Engineering.

Solvent hydrogenation of the 180°C-430°C liquids occurs in a conventional fixed bed catalytic hydrotreater. This restores the hydrogen donor capability of the solvent prior to being recycled as the coal slurrying medium. The catalytic reactor operates at approximately 12 300 kPa and the hydrotreated liquid product is fractionated to yield naphtha and solvent. The small quantity of hydrogenated naphtha is blended with the raw naphtha product from distillation. The hydrogenated solvent produced in excess of that required for liquefaction becomes a low sulphur mid-distillate product.

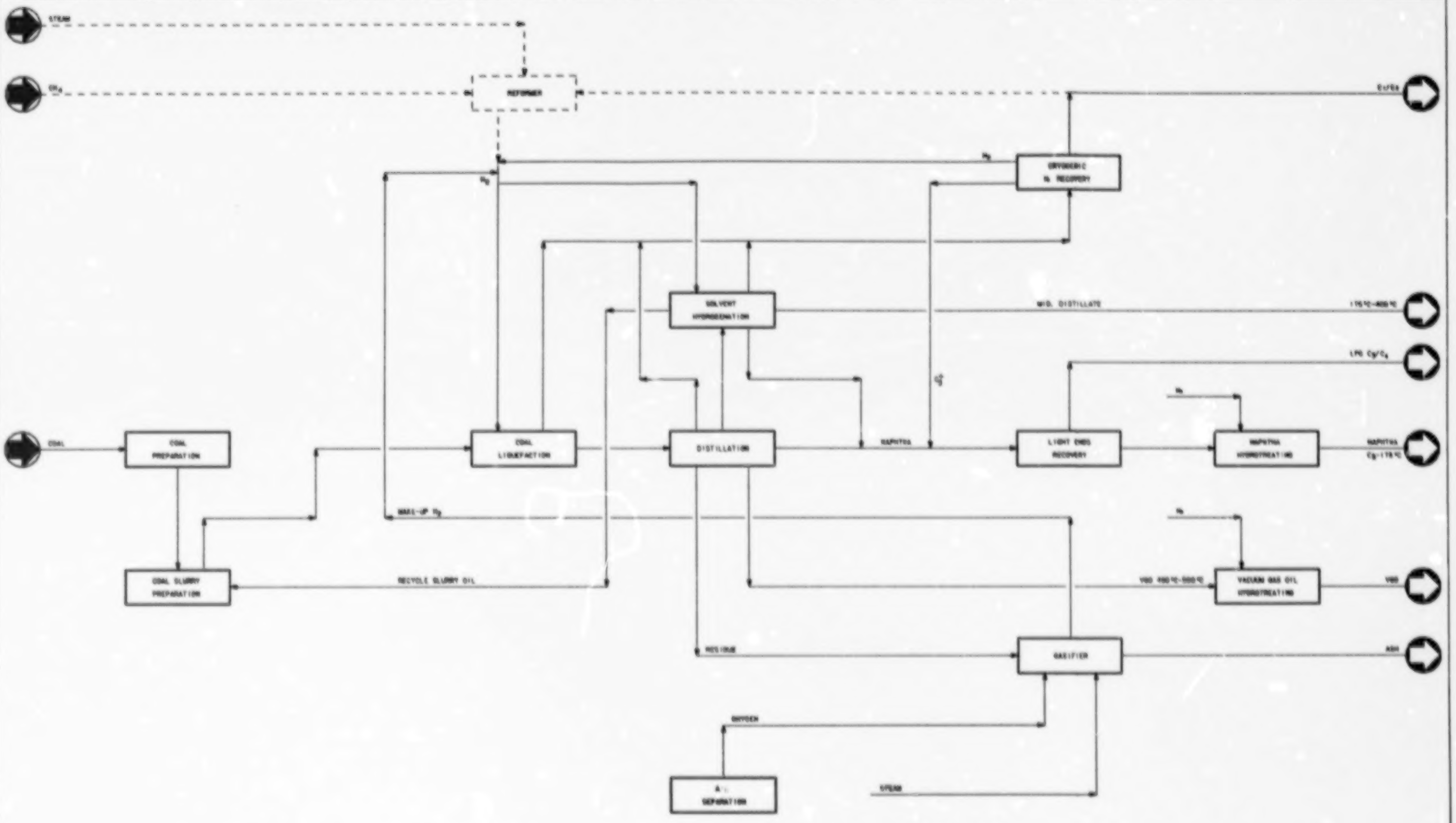
Both the naphtha and vacuum gas oil fractions are unstable as produced. Hydrotreatment is required in order to reduce the heteroatom concentration prior to storage or shipping. The naphtha could then be reformed using a conventional catalytic reforming process to yield a high octane gasoline. Severe hydrotreating of the vacuum gas oil is necessary to reduce the heteroatom content sufficiently to yield an acceptable industrial fuel. However special handling is required since it is incompatible with petroleum derived heavy distillate. Status of development of coal-derived liquid product upgrading processes is discussed in Section 4.7.

4.5.3 Process Development

As previously mentioned, Exxon conducted considerable research prior to and during the development of the EDS process. Consequently Exxon have extensive small scale experimental equipment varying from 100 cc batch units to a continuous 0.9 tonne per day integrated plant. In the mid 1970's design commenced on a 140 tonne per day pilot plant which was sponsored by a consortium including the U.S. Department of Energy and Electric Power Research Institute. This plant was started up in 1979 and operation since that time has been good. One year of testing Illinois No. 6 coal was completed after which the plant was modified to incorporate the "Bottoms Recycle" mode of operation. Testing with Wyodak coal is currently in progress and operation appears to be satisfactory apart from an occasional coking problem in the slurry preheat furnace. A 60 tonne/day flexicoking unit, originally intended to demonstrate the vacuum bottoms processing route, is no longer part of the EDS project.

4.5.4 Commercialization Plans

No commercialization plans have been revealed by Exxon as licensing terms and conditions have yet to be agreed upon by the program sponsors.


STEAM
CH_4
REFORMER
C_1/C_2
H_2
CRYOGENIC
H_2 RECOVERY
H_2
SOLVENT
HYDROGENATION
MID. DISTILLATE
175°C-400°C
LPG C_3/C_4
C_5^+
H_2
COAL
COAL
PREPARATION
COAL
LIQUEFACTION
DISTILLATION
NAPHTHA
LIGHT ENDS
RECOVERY
NAPHTHA
HYDROTREATING
NAPHTHA
C_5-175°C
MAKE-UP H_2
H_2
COAL SLURRY
PREPARATION
RECYCLE SLURRY OIL
VGO 400°C-500°C
VACUUM GAS OIL
HYDROTREATING
VGO
RESIDUE
GASIFIER
ASH
OXYGEN
AIR
SEPARATION
STEAM

Western Canada Coal Liquefaction Study
Prefeasibility Study
Simplified Block Flow Diagram
EDS Process

4.6 SRC-II

4.6.1 Introduction

The Pittsburgh and Midway Coal Company (a Gulf Oil subsidiary) have been developing Solvent Refined Coal (SRC) technology for almost 20 years. The original process, known as SRC-I, was conceived in order to produce a low sulphur, ashless feedstock for utility plants. It was intended to reduce atmospheric pollution by using this feedstock instead of high quality coal. This SRC-I process was subsequently modified to produce primarily liquid fuels by a process known as SRC-II.

In the SRC-II process the coal is first dissolved in a process derived recycle slurry. The dissolved coal is then hydrogenated at elevated temperature and pressure in the presence of hydrogen. The hydrogenation and hydrocracking reactions are promoted by the catalytic activity of the coal ash which is present both in the feed coal and in the recycle slurry.

4.6.2 Process Description

Raw coal is pulverized and dried then mixed with hot slurry recycled from the process. The slurry is pressurized to about 14 000 kPa and fed together with hydrogen through a fired heater to raise the temperature of the mixture to approximately 400°C. At this temperature the coal almost completely dissolves in the solvent and it is then hydrocracked in the reactor in the presence of additional hydrogen.

The reactor effluent is depressurized through a series of liquid-vapour separators whence it is ultimately separated into process gas, light hydrocarbon liquids and slurry. The process gas, consisting of hydrogen and gaseous hydrocarbons together with minor amounts of H_2S and CO_2 is cooled prior to passage through a conventional acid gas plant for removal of the H_2S and CO_2. The purified gas is then sent to a cryogenic plant for separation of the hydrocarbons. The purified hydrogen is recycled to the liquefaction step of the process while the hydrocarbons may also be used within the process or sold as byproducts.

All the light hydrocarbon liquids produced in the condensation steps are sent to an atmospheric distillation column. In this column the liquid is fractionated into naphtha (C_5-200°C), middle distillate (200-320°C) and atmospheric bottoms. This atmospheric residue stream is fed to a vacuum distillation column where it is used to strip the lighter components from a portion of the product slurry which is also fed to the vacuum column. The remainder of the product slurry is recycled for slurrying with the feed coal. Heavy distillate product is removed as vacuum tower sidestream while the overhead gas stream is sent to acid gas removal. The residue from the bottom of the vacuum tower would normally be gasified and processed to produce hydrogen for the process. Alternatively the residue may be used as a boiler fuel.

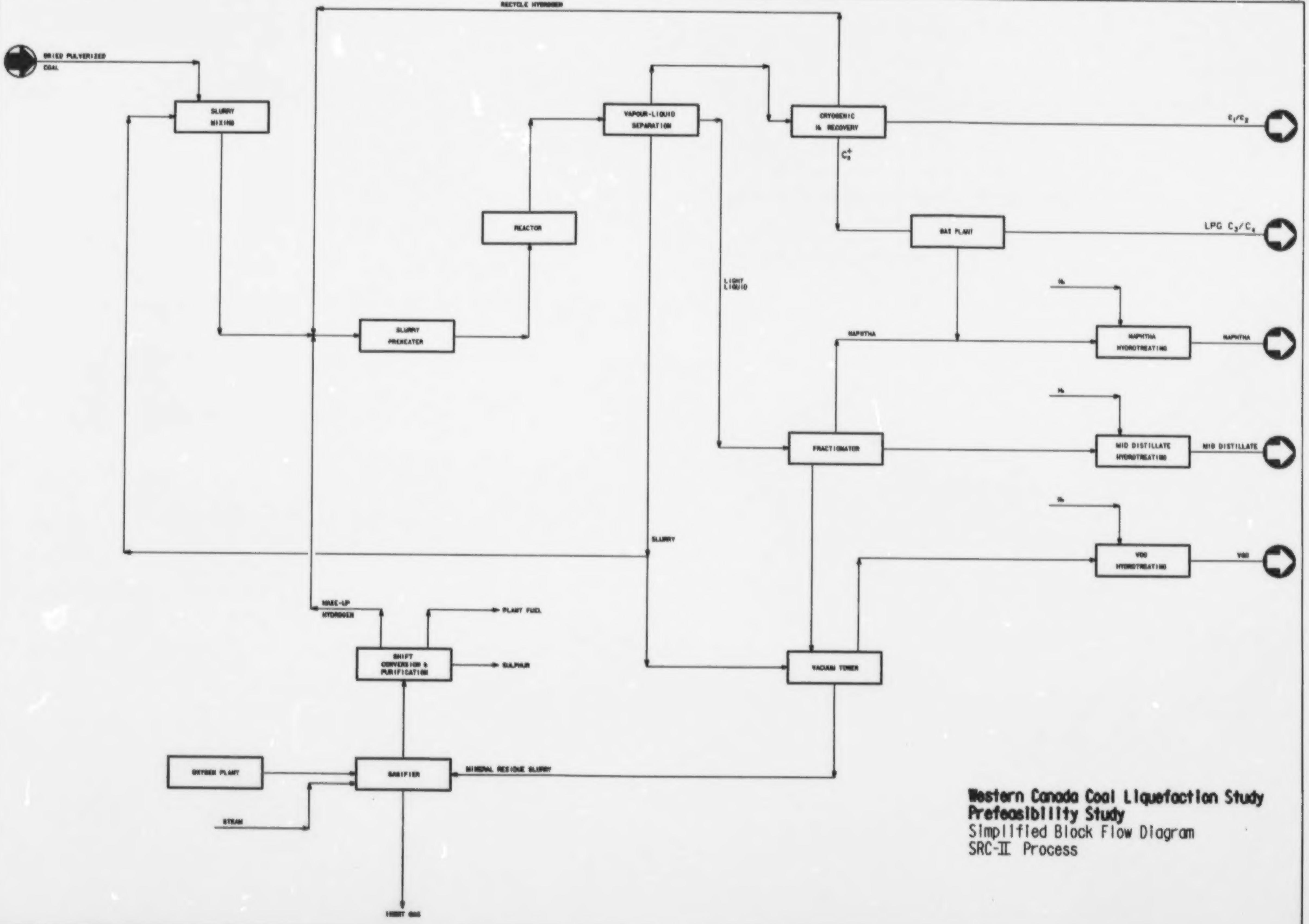
The liquid products, except the LPG, require substantial upgrading prior to use as finished products. Additional processing may vary from moderate hydrotreating to fluid catalytic cracking or hydrocracking depending on particular products required. Details on the status of development of upgrading options are given in Section 4.7.

4.6.3 Process Development

In 1974 a 45 tonne/day plant was completed at Fort Lewis, Washington, in order to pilot the SRC-I process. This plant was upgraded in 1977 to permit testing of the SRC-II process. Data collected from this plant and two other smaller test units were used to initiate the process design of a 5 400 tonne/day demonstration plant.

4.6.4 Commercialization Plans

The demonstration plant was to be built in Morgantown, West Virginia, but in mid 1981 the project was cancelled prior to the start of construction. No further commercialization plans have been announced. In fact, there appears to be no further committment by SRC International Inc. to commercialize the process and the Fort Lewis pilot plant has been mothballed.


RECYCLE HYDROGEN
DRIED PULVERIZED COAL
SLURRY MIXING
VAPOUR-LIQUID SEPARATION
CRYOGENIC H_2 RECOVERY
C_1/C_2
C_3^+
REACTOR
GAS PLANT
LPG C_3/C_4
LIGHT LIQUID
H_2
NAPHTHA
NAPHTHA HYDROTREATING
NAPHTHA
SLURRY PREHEATER
H_2
FRACTIONATOR
MID DISTILLATE HYDROTREATING
MID DISTILLATE
H_2
SLURRY
VGO HYDROTREATING
VGO
MAKE-UP HYDROGEN
PLANT FUEL
SHIFT CONVERSION & PURIFICATION
SULPHUR
VACUUM TOWER
OXYGEN PLANT
GASIFIER
MINERAL RESIDUE SLURRY
STEAM
INERT GAS

Western Canada Coal Liquefaction Study
Prefeasibility Study
Simplified Block Flow Diagram
SRC-II Process

4.7 SECONDARY PROCESSING OPTIONS

4.7.1 Introduction

Secondary processing options, essential to the successful development of direct coal liquefaction technology for the production of liquid transportation fuels, are in various stages of development or commercial readiness. These processes include raw liquid product upgrading options and hydrogen production via residue gasification.

Upgrading of raw coal-derived liquid products is necessary for the production of high quality synthetic crude that could be processed in a conventional petroleum refinery or for the production of specification liquid transportation fuels as the raw products contain high concentrations of heteroatoms (O, N, S) and are deficient in hydrogen. Process options include hydrotreating and catalytic reforming of the naphtha fraction and hydrotreating, hydrocracking and/or catalytic cracking of middle and heavy distillate fractions.

Hydrogen production via vacuum residue gasification is important as a source of hydrogen for the liquefaction complex and as a means of disposal of the residue, which has been identified as containing carcinogenic components. Only if a low price source of natural gas is available does hydrogen production via steam reforming of methane become attractive.

4.7.2 Raw Product Upgrading Options

Upgrading studies and pilot scale investigations have been carried out on coal-derived liquid fractions by a number of process developers in the United States and Europe, including Hydrocarbon Research Inc., UOP Inc., Mobil Oil Corporation, Chevron Research Company, Exxon Corporation, Ashland Oil Inc., Veba Oel AG, BASF Ludwigshafen, British Petroleum and others. These studies have included hydrotreating, hydrocracking, catalytic cracking, coking, catalytic reforming, partial oxidation and storage stability. Specification gasoline, jet fuel, diesel fuel, home heating oil and industrial turbine fuels have been produced in the pilot plants via the various options.

Coal-derived naphthas have been hydrotreated and catalytically reformed in research pilot plants to give high yields of 100-103 Research Octane Number (RON) gasoline containing 70-85 vol % aromatics. The naphthas were hydrotreated over commercial catalysts to reduce the sulphur and nitrogen contents to below 1 ppm. Operating conditions were relatively severe compared to those required to treat conventional petroleum naphthas. The hydrotreated naphthas were very responsive to catalytic reforming over commercial reforming catalysts (Pt-Re) and reforming proceeded at relatively mild conditions. Hydrogen yield was greatly in excess of the amount required for hydrotreating.

Middle distillates from direct liquefaction processes have been converted by hydrotreating, hydrocracking and fluid catalytic cracking (FCC). In general, hydrotreatment is required prior to either hydrocracking or FCC to reduce excessive heteroatom content, primarily oxygen and nitrogen, and to enhance processability. Hydrotreating alone has been shown to give high yields of environmentally acceptable home heating oil and No. 2 industrial fuel oil.

Low severity hydrocracking of hydrotreated middle distillate to specification jet fuel and diesel fuel has been reported by some process developers, notably Veba Oel and BASF. Hydrocracking of the middle distillate to gasoline and jet fuel has been investigated by some companies. Although the hydrocracking proceeded at operating conditions somewhat more severe than required for conventional petroleum distillates, high yields of high quality products have been reported. By reforming the hydrocracked naphtha, RON's in the order of 100 can be achieved. Premium jet fuel is produced, meeting smoke point and thermal stability specifications.

Heavy distillates have been converted in research pilot plants by hydrotreating, hydrocracking, FCC and coking to acceptable quality industrial or transportation fuels. Hydrotreating is required in all cases as the first processing step to reduce heteroatoms and to enhance processability by increasing hydrogen content. Hydrotreating alone has been shown to give acceptable quality industrial turbine fuel.

Pilot plant studies, particularly by Veba Oel, BASF and Chevron Research, have shown that specification naphtha and diesel fuel can be produced from the heavy distillate fraction via two-stage hydrocracking. Operating conditions can be varied to maximize the yield of either naphtha or diesel fuel. FCC research pilot plant tests on hydrotreated heavy distillate (or middle + heavy distillate fractions) have been encouraging, resulting in high yields of gasoline and cycle oil. The FCC gasoline product octane number ranged from 92-98 RON, whereas the cycle oil could be used as a diesel fuel or home heating oil blend stock.

While upgrading of coal-derived liquid hydrocarbon fractions to transportation fuels, home heating oil or industrial turbine fuels has not been demonstrated on a commercial scale, the process developers are confident that research scale pilot plant data can be scaled up to commercial size -- as is common practice with petroleum -derived hydrocarbon streams. Pilot plant tests would have to be conducted on liquid fractions from the specific coal under consideration to determine optimum operating conditions and catalyst deactivation rates, to provide a basis for process design of the commercial unit and to establish process guarantees.

4.7.3 Residue Gasification

Gasification of high ash H-Coal residues obtained from the liquefaction of Illinois No. 6 and Wyodak coals has been successfully demonstrated in large scale pilot plant equipment at the Montebello Research Laboratory of Texaco Inc. Approximately 35 tonne of each residue, containing up to 20 percent ash, were gasified at feed rates up to 0.4 tonne per hour, at a pressure of 24 atmospheres.

The synthesis gas produced consisted of over 92 volume percent carbon monoxide and hydrogen on a dry basis, at an average ratio of 1.35 moles CO per mole of H_2 (or a molecular H/C ratio of 1.48/1). Conversion of the synthesis gas to high purity hydrogen was not demonstrated as commercial processes exist for this processing step.

Overall carbon conversion efficiency varied between 97 and 99 percent for both residues. Cold gas efficiencies averaged between 83 and 84 percent. Steam and oxygen requirements were comparable to those used with heavy oil gasification. Operating temperature in the gasifier ranged from 1280°C to 1410°C.

No major operating problems were encountered during the pilot plant operation which included feeding of the molten residue to the gasifier, gasification, syngas quenching and scrubbing, slag removal and soot recovery. No evidence of refractory damage or serious burner tip erosion was observed in the 200 hours of operation.

Texaco report that sufficient data were obtained to verify existing computer correlations and to design a commercial plant for the efficient gasification of these two, or other similar, coal liquefaction residues.

It is understood that coal liquefaction residues have been tested in other gasifiers, including the Shell-Koppers entrained bed gasifier the Saarberg/Otto slag bath gasifier. However, detailed test results are not available. In addition, Ruhrkohle/Veba are planning a major coal liquefaction residue gasification test in the Texaco Oberhausen-Holten entrained bed gasifier, as part of their 200 tonne per day coal liquefaction pilot plant program. Ruhrkohle AG and Ruhrchemie AG are partners in the Texaco gasifier program.

Additional evidence that the Texaco Partial Oxidation Process is considered to be commercially ready is the apparent preference for this gasification process in the proposed Breckenridge, Kentucky, H-Coal facility. Gasification of the residue is considered to be a key step in the project.

5.0 COAL TESTING PROGRAM

5.1 OVERVIEW

The feasibility of this proposed coal conversion project primarily depends upon the liquefaction potential of the two nominated coals. Ideally under the operating conditions of the direct liquefaction processes selected, 100 wt% maceral content of the coal feed would be converted into liquids in the LPG/gasoline/mid-distillate boiling range. The liquids obtained in this ideal situation would also contain sufficiently few impurities that they would require little or no further treatment to yield finished products. In practice, this ideal situation is never achieved and more commonly the combination of a good process and feed coal would result in approximately 60% by weight of the coal macerals being converted into liquids. These liquid products are in the LPG to heavy distillate range and the higher the heavy distillate or pitch-type liquid yield, the less attractive is the process and/or the feed coal.

The reactivity of the Greenhills and Monkman coals were tested under conditions similar to those used by the process developers. Under the limited test conditions, it is recognized that true simulation of any particular process was not possible. However, attempts were made to recreate the conditions used by the individual process developers; these attempts have been termed as 'model', 'modelling', 'simulate' or 'simulation', where appropriate, throughout the text.

The determination of the liquefaction potential of the B.C. coals was conducted in two stages. In the first stage, microautoclave tests were performed by B.C. Research to 'model' four direct liquefaction processes. The two catalytic processes 'modelled' were the H-Coal process and the Ruhrkohle/Veba (German) process. SRC-II and EDS were the two non-catalytic processes 'simulated'. It was not possible to 'simulate' the NCB/BP LSE process because of the limited scope of the work. To obtain a direct comparison of the liquefaction potential of the B.C. coals, Illinois No. 6 was used as a reference coal.

The Ruhrkohle/Veba process was 'simulated' using a coal-derived recycle solvent obtained from Bergbau-Forschung in Germany and Red Mud (Fe_2O_3) catalyst. Hydrogenated anthracene oil was used as the solvent for 'simulating' the other three processes. The catalyst used in 'modelling' the H-Coal process was HDS 1442A, a Cobalt-Molybdenum catalyst used by HRI.

The B.C. coals were found to be far less reactive than Illinois No. 6 coal. Conversions in the range of 40-68 percent were achieved (compared to 54-80 percent for Illinois No.6), but the distillate oil fractions were only in the range of 25-50 percent.

The distillate oil yields were in the order of 10 percent higher in catalytic processes under similar conditions. On comparison of the two catalytic processes, the coal-derived solvent and Red Mud catalyst (German process) showed about 10 percent better yields. However, one test under similar operating conditions, with hydrogenated anthracene oil and Red Mud catalyst gave 10-18 percent lower distillate oil yield, indicating a very strong dependency on the nature of the solvent. In earlier tests with raw anthracene oil as solvent, the yields were much lower than with hydrogenated anthracene oil.

Having established the better performance of the catalytic processes, the basis for tests on the SBRG continuous process development unit were established. The two catalytic processes to be 'simulated' were H-Coal and Ruhrkohle/Veba. Hydrogenated anthracene oil was the common solvent used with both HDS and Red Mud catalysts.

The test using H-Coal catalyst was performed at the temperature and pressure normally used by the process licensor. Due to equipment limitations, the operating temperature for 'simulating' the Ruhrkohle/Veba process was about 30-35°C lower than the recommended temperature.

Since a true coal-derived recycle solvent was not available, actual distillate yields could not be determined. Instead, a minimum and maximum range of distillate yields was calculated. This range was based on assumptions made on the solvent behaviour.

Using these assumptions, the distillate yields from the Monkman coal 'simulated' in the H-Coal process were between 19% and 34%. The Ruhrkohle/Veba 'simulation' with the Greenhills coal yielded between 18% and 34% distillates.

Again, high liquid yields were obtained, most of which was a heavy oil fraction or pitch. The absence of a true recycle solvent and, in the case of the Ruhrkohle/Veba process, lower than desired operating temperature is recognized as affecting distillate yields.

Both the SBRG tests were conducted under predetermined conditions and not experimentally optimized conditions. This was also recognized by the process developers who suggested that independent tests be performed by themselves, which would simulate recycle solvent operation to determine the liquefaction potential of the coals at optimized conditions.

HRI performed a set of microautoclave tests which, they claimed, would provide data that could be correlated against their data bank and would enable prediction of the obtainable distillate yields. HRI predicted that, with recycle of the vacuum bottoms residue, 43% distillate yield could be obtained by the H-Coal process. The actual split of the distillate oil fractions was based on assumptions made on the possible solvent behaviour--as in the SBRG tests. HRI predicted two extreme values of the distillate oil split:

1.8%, 5.8% and 35.4% as one extreme, and

3.7%, 19.9% and 19.4% as the other extreme

for the light distillate, middle distillate and heavy distillate fractions, respectively.

On reviewing the HRI report, it became evident that the results of the microautoclave tests conducted by BCR and HRI are very similar, within acceptable experimental accuracy (about ±5%).

Tests are also being conducted by Veba Oel in West Germany. The results are not expected to be available in time to be included in this report.

5.2 GENERAL

One of the major objectives set forth in the Prefeasibility Study was to select which of the two nominated B.C. Coals were best suited for liquefaction. The selection was to be made by performing laboratory tests on samples of the two coals selected by the Western Canada Coal Liquefaction Study Consortium. B.C. Research (BCR) was nominated to perform the first stage of the testing program which was a series of microautoclave tests on the Greenhills, Monkman and Illinois No. 6 coals. Illinois No. 6 coal was tested as a reference coal for comparison purposes.

Based on the results of the microautoclave tests, optimum operating parameters would be established for the liquefaction of the two nominated coals. These parameters would then be used as a basis for the determination of operating conditions for continuous tests to be performed by Sandwell Beak Research Group (SBRG). The continuous tests results were expected to indicate the maximum distillate yield and probable product slate for each coal.

Microautoclave tests were performed by HRI to simulate recycle solvent operation of the H-Coal process with the nominated B.C. coals. HRI predicted that their data would simulate the H-Coal process and would aid in selection of the better candidate coal.

5.3 COAL SAMPLING, HANDLING AND PREPARATION

The selection of one of the two coals and subsequent selection of a process is dependent, to a large extent, on the results of tests carried out in Canada and further testing by selected process developers. The validity of the test results and the recommendations made are highly dependent on the quality of the coal samples provided.

To ensure obtaining a representative sample, B.C. Coal undertook to supervise the sampling of the Greenhills coal. Two drums of minus one inch sample were taken from seam number 1 and one drum each.from seam numbers 7 and 10. The sealed drums were then shipped to B.C. Research facilities in Vancouver. A small sample was taken from each drum and mixed in the proportions recommended by the B.C. coal, 50% from seam number 1 and 25% each from seam numbers 7 and 10. After taking the quantity of coal required for their tests, B.C. Research shipped the bulk of the Greenhills coal in sealed containers to Ontario Research Foundation (ORF) in Toronto for blending and grinding.

The Monkman coal sample was provided by Petro-Canada from the Birtley Laboratory in Calgary. Although a fresh sample was not taken, a representative sample from storage was provided. Analysis carried out by BCR on this sample showed no effects of oxidation. The BCR coal analysis is included in their report, included in Volume III of this report. Sufficient quantities of this sample were sent to B.C. Research and the bulk, nearly 600 kg of the minus one inch mesh, was shipped in sealed containers to ORF in Toronto.

ORF was contracted to blend and grind both minus one inch coal samples to pass minus 100 mesh size. This was the size required by SBRG for the continuous tests. During the blending and grinding operation, care was taken to minimize the risk of oxidation. Both the blended and ground coal samples were placed in appropriately marked sealed containers in a CO_2 inert atmosphere.

5.4 SAMPLING AND PREPARATION OF RED MUD CATALYST

The German process, one of the processes considered and proposed for 'simulation' in the test program, uses Red Mud (RM) as a catalyst. Red Mud, rich in Fe_2O_3, is a waste product from the Aluminum industry using bauxite as a feedstock.

British Columbia has no available source of Red Mud whereas ALCAN Industries in Arvida, Quebec, produce substantial quantities as waste product. At Kilborn's request, ALCAN provided samples of the Red Mud slurry to B.C. Research in Vancouver and SBRG in Toronto.

The Red Mud slurry was dried, ground and sampled for the test program. Because of the substantial quantities required for the continuous tests, the drying and grinding of the Red Mud was completed by ORF. The few grams required by B.C. Research was prepared in-house. The analysis of the RM catalyst is shown in Table 5 of the SBRG report attached in Volume III of this report.

5.5 B.C. RESEARCH MICROAUTOCLAVE TESTS

5.5.1 Objectives of the Tests

These bench scale tubing bomb tests were performed to provide initial data on conversion and distillate yield under selected conditions for the two nominated B.C. coals and Illinois No. 6, the reference coal. Tubing bomb tests are normally used only to compare coal reactivities and identify process(es) having the best potential for liquefaction. The specific yield data for the microautoclave experiment should be regarded only as a qualitative indication of a pilot scale or commercial system performance.

5.5.2 Definition of Test Program

The test program was designed to obtain the maximum relevant information with the minimum of tests, dictated by budget limitations. The test program was designed to 'model' the H-Coal, Ruhrkohle/Veba, EDS and SRC-II processes.

The H-Coal process was 'simulated' using prehydrogenated anthracene oil as solvent and HDS 1442A catalyst (a Co/Mo based catalyst used by HRI). Obtaining a true coal-derived recycle solvent from HRI was not possible. The 'simulation' of the Ruhrkohle/Veba process was achieved by using a coal-derived recycle solvent obtained from Bergbau-Forschung (BF) and the Red Mud catalyst. The non-catalytic SRC-II and EDS processes were 'simulated' by reacting the coal with prehydrogenated anthracene oil.

The preference of using prehydrogenated anthracene oil over raw anthracene oil was dictated by preliminary tests performed by BCR. The inactivity of the raw anthracene oil is attributed to its low hydrogen content. Being deficient in hydrogen, the anthracene oil preferentially tends to utilize a large quantity of hydrogen in the autoclave reaction, leaving very little hydrogen for coal conversion. Prehydrogenating the anthracene oil would reduce its reactivity with hydrogen and allow the coal to react with more hydrogen, hence a better conversion.

For the non-catalytic processes, the parameters varied were the solvent-to-coal ratio and the reactor operating temperature. The variables investigated for the catalytic processes were the solvent-to-coal ratio, reactor operating temperature and the catalyst concentration. Residence time was not considered a determining factor.

5.5.3 Test Parameters

In all the tests performed, the hydrogen pressure in the cold reactor was set at 7000 kPa. The pressure during the reaction was not monitored. The heating rate of the reactor was 3-4 minutes and the slurry residence time was set at 25 minutes. These three parameters were fixed based on BCR's previous experience. Although the slurry residence time for almost all of the test runs was kept at 25 minutes, some tests were performed at 10 minutes and 60 minutes to determine the effect of varying residence time.

For the non-catalytic tests, the solvent-to-coal ratio was varied from 2:1 to 3:1. The effect of temperature at 420°C, 440°C, 460°C and 480°C was also determined.

For the catalytic processes, in addition to the solvent-to-coal ratio and temperature variation, two levels of catalyst concentrations were studied. For the HDS 1442A catalyst, the High Level corresponded to 2.0 wt% of coal, as elemental molybdenum, with the Low Level being equivalent to 0.2 wt% of coal charge, as elemental Mo. The High level for the Red Mud catalyst corresponded to elemental iron as 5.0 wt% of the coal charge, the Low Level being equivalent to 2.0 wt% as Fe.

Additional tests were performed to determine the effect of the solvents alone with varying catalyst concentrations and at different temperatures. This would give an indicaton of the pitch formed and the hydrogen consumption by the solvent. The temperature range and catalyst concentrations were similar to those described above.

To establish the effect of ash content of the coals, tests were carried out on washed coal samples. The degree of washing is included in the BCR report. The results were to be compared with the run-of-mine coal samples under similar operating conditions.

5.5.4 Test Result Determination and Presentation

After completion of the experiment, the contents of the tubing bombs are washed with CS_2 solvent. The liquid products (distillate and heavier fractions) are recovered in the CS_2 solvent, leaving the unreacted coal, mineral content and catalyst (if used) as residue. The amount of unreacted coal is easily determined knowing the mineral and catalyst contents (assuming the quantity of the catalyst does not change).

The pitch (heavy fraction) content in the converted liquid is determined by thermogravimetric analysis of a small sample of the CS_2 solubles, while the different fractions of the distillate products are determined by simulated distillation of the CS_2 soluble products.

The amount of gas products and the hydrogen consumption are determined from the gas analyses and the known volume of the reactor and gas lines.

The presentation of the results include the operating conditions, gas formation and liquid conversion, hydrogen consumption, residue in CS_2 solubles, and, where calculated, minimum and maximum distillate yields. The BCR report is included in Volume III of this report.

5.5.5 Discussion of BCR Microautoclave Test Results

The liquefaction potential of a coal is determined by its conversion to distillate oils. In the microautoclave tests, the determination of the distillable fractions is based upon the estimate of pitch in the CS_2 solubles. The CS_2 soluble pitch originates from two sources. When coal is hydrogenated one of the fractions formed is the heavy residue. The solvent used, the other source, has an inherent pitch content which may or may not reduce during the hydrogenation process.

The quantity of inherent pitch in the solvents was determined. Hydrogenation of the solvents alone, under conditions at which the coals were tested (presence of catalyst, operating temperature, residence time) reduced the solvent pitch. However, in the presence of coal and under similar operating conditions, the solvent behaviour may not necessarily be the same due to interaction between the coal and solvent.

Although the inherent pitch content of the solvents may change due to hydrogenation in the presence of coal and catalyst (if a catalysed reaction), quantifying the inherent pitch content facilitates making certain assumptions to define the distillate yields from a liquefaction process. The possible minimum and maximum distillate yield values are computed assuming the origin of the pitch source. The minimum distillate yield is calculated on the assumption that all the CS_2 soluble pitch was derived from the coal. This hypothesis assumes complete elimination of solvent pitch during reaction and that all the pitch is coal-derived. The maximum distillate yield is calculated assuming that the solvent pitch content does not reduce, and the additional pitch formed is due to liquefaction of coal, i.e. minimum conversion of coal to pitch.

This procedure merely indicates the possible range of yields that can be theoretically obtained. The actual yields could be anywhere in between the two extreme values.

Within the experimental confines the distillate yields from both the B.C. coals were very similar for all the operating conditions. However, the Greenhills coal consistently showed a 1-5 percent higher yield than the Monkman coal. In contrast, Illinois No. 6, the reference coal, was 10-30 percent more reactive than both the B.C. coals, particularly with HDS Catalyst. This can be observed from key selected results listed in Table 5.1. The detailed BCR results are attached in Volume III.

From Table 5.1, the BF solvent/RM catalyst combination favours conversion of B.C. coals. Although relatively high conversions are obtained, the distillate yields are significantly low. For this solvent/catalyst combination, 460°C appears to be the optimum temperature. Substantially increased yields are obtained at 460°C compared to 440°C. Further increasing the temperature to 480°C increases the gas production but slightly lowers the distillate oil yield.

Under similar non-catalytic operating conditions, increased CS_2 solubles were observed with the BF solvent compared to the HAO solvent. The BF solvent has a higher hydrogen content than the HAO having been recycled several times. Hence, it acts as a better hydrogen donor.

The addition of the Low concentration of catalyst to either of the solvents increases the CS_2 solubles substantially. Further increasing the catalyst concentration to the High Level showed moderately additional increased yields. This was true for both the BF/Red Mud and HAO/HDS combinations. Increasing the residence time from 25 minutes to 60 minutes showed marginal increases in CS_2 solubles. This was consistent wth previous work that BCR had undertaken. Hence the residence time for all the tests was kept at 25 minutes.

TABLE 5.1 *

Coal	Solvent	Solvent/Coal Ratio	Catalyst Concentration	Temp °C	Residence Time	Conversion wt % DAF Coal CS_2 Soluble	Distillate Yields wt% DAF Coal	
							Min.	Max.
GH	BF	2:1	HIRM	460	25	67.8	42.9	50.8
MKM	BF	2:1	HIRM	460	25	66.3	41.0	49.3
I6	BF	2:1	HIRM	460	25	81.1	44.5	52.3
GH	BF	2:1	HIRM	480	25	70.2	41.2	49.1
MKM	BF	2:1	HIRM	480	25	63.7	33.5	41.7
I6	BF	2:1	HIRM	480	25	79.0	54.2	61.9
GH	HAO	2:1	HIHDS	440	25	54.2	30.4	31.8
MKM	HAO	2:1	HIHDS	440	25	55.1	31.6	32.9
I6	HAO	2:1	HIHDS	440	25	77.4	61.1	62.3
GH	HAO	2:1	HIRM	460	25	58.0	31.2	32.4

NOTE:

BF = Bergbau-Forschung Solvent
HAO = Hydrogenated Anthracene Oil Solvent
GH = Greenhills Coal
MKM = Monkman Coal
I6 = Illinois No. 6 Coal
HIRM = High Concentration Red Mud Catalyst
HIHDS = High Concentration HDS Catalyst

* Extracted from BCR report.

Tests on clean coals under similar operating conditions as the run-of-mine coals did not indicate any increase in conversion. An insufficient number of tests prevented any conclusive inferences to be drawn.

One test run was performed to determine the influence of the solvent under otherwise optimum conditions. Greenhills coal was tested with HIRM catalyst concentration at 460°C and 25 minutes residence time using a 2:1 HAO-to-coal ratio. Compared to results with BF solvent, the overall conversion and the liquid yield is considerably reduced. With all other parameters being the same, the quality of the solvent appears to have a considerable effect on coal conversion. The HAO, being hydrogenated only once, may crack rather than participate in the hydrogenation process whereas BF solvent is a truly recycle solvent and is not cracked.

In conclusion, under the conditions at which the tests were performed, the B.C. coals react favourably and produce increased yields with the German technology 'modelled' process, e.g. BF solvent and Red Mud catalyst, lower yields with the H-Coal 'modelled' process (HAO/HDS catalyst) and very low yields with non-catalytic processes. However, it should be recognized that the substantially increased yields in these tests with the German process may have been induced by the very high catalyst concentrations and a better hydrogen donor solvent. The Low catalyst concentration corresponded to about 10.5 wt% on MAF coal and the High concentration was 25% for the Greenhills coal. The German process developers typically use 5 wt% catalyst on MAF coal.

The 'modelling' of the H-Coal process (last set of data on Table 5.1) shows relatively low distillate yields for the B.C. coals in comparison to Illinois No. 6. In this case too, the High catalyst concentration corresponds to about 26% catalyst weight on MAF coal.

The H-Coal 'modelled' process was not tested at temperatures above 440°C as coke formation increases with higher temperatures. Red Mud acts as a stronger coking inhibitor than the HDS catalyst and functions better at higher temperatures.

5.6 PDU TESTS

5.6.1 Definition and Scope and the Test Program

The continuous process development unit (PDU) tests were conducted by SBRG to demonstrate the viability of continuous liquefaction of the nominated B.C. coals. It was hoped that adequate data would be obtained from the PDU tests which, along the microautoclave test results, would permit the process developers to predict the performance of their processes. It was also hoped that the completed experimental program (both phases) would enable selection of the B.C. coal and process best suited for liquefaction to produce synthetic crude.

The operating parameters for the PDU test program were to be determined from the tubing bomb test results and SBRG's previous experience.

5.6.2 Operating Conditions

With the BF coal-derived solvent, reasonable yields were obtained in the microautoclave tests. To process a coal in a PDU, very large quantities of the coal-derived solvent would be required. Budget and time restrictions prevented production of sufficient quantities of recycle solvent to process the coals. Consequently, raw anthracene oil was catalytically hydrogenated in the continuous unit to increase its hydrogen donor capability. (In the microautoclave tests, higher yields were obtained with hydrogenated anthracene oil than with raw anthracene oil).

The two experiments to be performed were to 'simulate' the catalytic processes. The two catalytic processes 'simulated' (within experimental/equipment limitations) were the German process and the H-Coal process. The BCR tests did not favour non-catalytic processes.

In the tubing bomb tests the Greenhills coal gave marginally better yields than the Monkman coal. The slightly improved yields are consistent with the higher volatile matter in the Greenhills coal. Also, the Red Mud catalyst showed a definite trend of being a better catalyst than the HDS 1442A catalyst. In view of this, the combination for the test program chosen was:

Greenhills coal with Red Mud catalyst
Monkman coal with HDS 1442A catalyst.

This combination was based on the test results obtained and the researchers' judgements.

The H-Coal process was 'simulated' at a reactor pressure of 20 800 kPa and at a reactor temperature of 440°C. The catalyst content was set at 2.0% of HDS catalyst on MAF coal. This was thought to be in line with H-Coal catalyst levels.

Ideally, the Ruhrkohle/Veba process should have been 'simulated' at a minimum reactor temperature of 460°C. This was the temperature the microautoclave tests predicted to give the maximum yields, although, the Bergbau-Forschung PDU tests are performed at temperatures in the order of 475°C. At the time, SBRG was reluctant to operate at a reactor temperature beyond 440°C due to equipment temperature limitations. The hydrogen pressure was set at 26 300 kPa, consistent with Ruhrkohle's recommendation with pure hydrogen make-up. The Red Mud catalyst concentration was set at 1.5% Fe_2O_3 on MAF coal. This is the recommended concentration for use in the German process.

5.6.3 Results and Discussion

The solvent used in a coal liquefaction process has a significant effect in the coal conversion. In a once through coal liquefaction process, the solvent itself undergoes change, the extent of which is not known unless specific methods are designed to determine these changes. These methods are difficult, expensive and time consuming. In contrast, in a continuous recycle system, the solvent has been through several hydrogenation steps and becomes stable and does not change its characteristics. Hence any liquid obtained, other than the solvent, is coal-derived.

In the SBRG liquefaction tests, the coal conversion is determined by dissolving the liquefaction products in THF solvent. The insolubles comprise unreacted coal, ash and catalyst. The ash and catalyst content are known; therefore the amount of converted coal is calculated. The liquid yield obtained constitutes the solvent (hydrogenated anthracene oil) and the coal derived liquid. To calculate the coal derived liquid yields, assumptions are made to define the solvent content.

The first assumption is based on the hypothesis that the solvent does not change in characteristics in the presence of coal and catalyst at the test conditions. Hence, the total liquid product less the solvent is coal-derived liquid. In the second case, all the light gases, i.e. C_1's, C_2's and C_3's are assumed to be formed as a consequence of the cracking of the solvent. Hence, the total liquid contains less solvent than was input, since some of the solvent has cracked and formed light gases.

The results of the two experiments on the Monkman and Greenhills coals are shown in Tables 5.2 and 5.3. Conversions, total liquid yields, distillate yields and gas yields are presented. The distillate yields in Table 5.2 are calculated on the first assumption i.e. no change in solvent characteristics. Table 5.3 shows the distillate yields based on total gas formation via cracking of the solvent.

For the Monkman/HDS combination, the minimum and maximum distillate yields obtained were 19% and 34% respectively. The Greenhills/Red Mud combination produced 18% and 34% distillate yields.

The two assumptions define the extremes of coal-derived liquid yields that can be calculated. The first assumption would be accurate for a true recycle solvent. However, this assumption defines the lowest possible yield that can be obtained. The second assumption determines the yields on the higher end of the scale. The assumption that all the gases are formed from the solvent cracking is unrealistic. This is evident from Table 5.3 in which the liquid yields, in both experiments, are higher than the conversion and negative gas yields are obtained. In effect, the actual yield could be anywhere in between the two extreme values. Both are very low compared to the observed microautoclave test results.

The Greenhills/Red Mud combination gave slightly poorer yields than the Monkman/HDS test. This was probably due to the lower than optimum reactor temperature. Red Mud appears to give better yields at higher temperatures.

The analysis of the liquid products from both the experiments indicated that the bulk of the liquids from these coals, about 97%, constitute the 200°C+ heavy fraction. The distribution of the distillation cuts from both the coal liquids, extracted from Tables 9 and 10 of the SBRG report, is shown in Table 5.4. This distribution is similar, within the accuracy of the experimental procedure, for both coal liquids. The high concentration of the heavy liquids from these coals are not easily hydrocracked or are resistant to hydrocracking.

The detailed SBRG report is included in Volume III of the report.

TABLE 5.2 *

Assumption: No Solvent Change

% MAF Coal	Monkman	Greenhills
Conversion	82.8	76.6
Liquid Yield	68.2	65.6
Gas Yield	12.0	9.8
Distillate Yield	19.0	17.7

TABLE 5.3 *

Assumption: All C_1, C_2 and C_3 Formation Via Solvent Cracking

% MAF Coal	Monkman	Greenhills
Conversion	82.8	76.6
Liquid Yield	83.3	79.3
Gas Yield	-3.1	-3.9
Distillate Yield	34.04	33.84

* Extracted from Table 8 of the SBRG report.

TABLE 5.4*

ANALYSIS OF COAL LIQUID PRODUCTS

Monkman Coal/HDS Catalyst

<200°C	2.8%
200-325°C	47.5%
325-475°C	22.8%
475°C+	26.7%

Greenhills/RM Catalyst

<200°C	3.2%
200-325°C	45.3%
325-480°C	22.3%
480°C+	29.2%

* Extracted from Tables 9 and 10 of the SBRG report.

5.7 HRI MICROAUTOCLAVE TESTS

5.7.1 Objectives of the HRI Test Program

The low distillate yields predicted in the SBRG continuous test program 'simulating' the H-Coal process were regarded by HRI as not being truly representative of the H-Coal process. HRI claimed that the absence of a recycle solvent, testing at non-optimized process conditions and the inability of the SBRG test facility to simulate an ebullated bed reactor did not fully demonstrate the H-Coal process capability and, hence, predicted low yields. HRI claimed that their microautoclave tests would simulate the H-Coal process more accurately than the SBRG test program.

5.7.2 HRI Microautoclave Test Procedure

The microautoclave test procedure employed by HRI was similar to that used by B.C. Research. One major difference is that BCR used CS_2 solvent to wash the contents of the reactor, whereas HRI used THF and toluene. The solubility range of CS_2 falls midrange between that of THF and toluene. The method of calculating coal conversion is also similar.

5.7.3 Test Parameters and Methodology

In all the tests performed, the hydrogen pressure was held at 13 900 kPa. The temperatures studied ranged from 400°C to 455°C and the slurry residence times were 10, 15 or 30 minutes.

The tests were performed in two stages. In the first stage the coal, in the presence of the solvent, was subjected to heat treatment at different temperatures for different residence times in the absence of a catalyst. In the second stage the catalyst was added in a basket to the products of the first stage test and heated to the specified temperature for the required residence time. The products were then analyzed. From the results of the above procedure, an optimized set of conditions (temperature and residence time) were determined.

In order to approximate the recycle slurry oil used by the continuous process, a two-pass test procedure was followed. In the first pass, the two-stage procedure, described above, was duplicated. To the material produced in the first pass fresh coal was added and the two-stage procedure repeated. The products were then removed and analyzed.

In all the first pass operations, hydrogenated anthracene oil in a 2:1 solvent-to-coal ratio was used. The second pass operations used an 8:1 solvent-to-coal ratio, the solvent being the products of the first pass procedure. This method, according to HRI, simulates vacuum bottoms recycle and the use of middle and heavy distillate fractions as recycle slurry oil.

The catalyst charge in all the tests was 0.5 g catalyst per gram of solids.

The detailed tests were performed on the Greenhills coal sample only. The mean reflective indices indicate that the Greenhills coal is slightly lower in rank than the Monkman coal, although the rank differential is not large. Also, the Greenhills coal contains higher reactive maceral content than the Monkman coal. HRI felt that a detailed analysis of the Greenhills coal sample would enable them to predict results for the Monkman coal, within reasonable accuracy. Limited testing of the Monkman coal was performed to verify the above assumption.

5.7.4 Discussion of the HRI Microautoclave Tests

The results of the first stage tests indicate that the conversion of coal to THF soluble material increases with increasing temperature and residence time. However, to obtain the maximum conversion with the two-stage procedure, the first stage parameters chosen were 455°C and 10 minutes residence time. At these conditions a 60% conversion of coal to THF soluble material was obtained.

The second stage was optimized and the optimum conditions were found to be 455°C and 10 minutes residence time. Operation at these conditions provided yields of 72% THF conversion and 67% toluene conversion with hydrogenated anthracene oil solvent.

The first stage tests, except for the optimization, is identical to the non-catalytic tests performed by BCR. In addition, the second stage tests are similar to the BCR catalytic tests. Although no direct comparison of the results of the two laboratory programs can be made due to different operating parameters, results of tests performed under similar conditions are compared.

A BCR non-catalytic test with a 2:1 HAO solvent-to-Greenhills coal ratio at 460°C and 25 minutes residence time yielded 41.5% CS_2 solubles, compared to 60% THF solubles in a similar HRI test at 455°C and 10 minutes residence time. Part of the difference is explained by the fact that THF is a better solvent than CS_2, resulting in a higher measured conversion. Furthermore, the hydrogen content of HAO used in the BCR test was 6.88%, compared to 7.58% for the HAO used in the HRI tests -- an important parameter as it was demonstrated in the BCR tests that coal conversion is highly dependent upon the slurry solvent characteristics. A similar BCR test with BF solvent yielded 47% CS_2 solubles. The hydrogen content of the BF solvent was 8.46%. Considering the nature of the two extraction solvents (THF and CS_2) and allowing for experimental accuracies, the difference between the HRI and BCR results is not great, recognising that the HRI tests were performed under optimized conditions.

Results of the BCR catalytic tests with Greenhills coal and HAO at a 2:1 solvent-to-coal ratio, using HDS 1442A catalyst and at 440°C and 25 minutes residence time yielded 54% CS_2 solubles. This relatively low conversion, compared to 72% THF conversion and 67% toluene conversion by the HRI tests can be attributed to several reasons, including the quality of the slurry solvent, lower reaction temperature, short residence time and non-optimized single stage operation.

By comparison, another BCR test with Greenhills coal, using BF solvent and RM catalyst, at 460°C and 25 minutes residence time yielded 68% CS_2 solubles, almost as high a conversion as obtained by HRI.

The results of the HRI tests to simulate the continuous H-Coal process under optimized conditions yielded a 90% THF and 70.4% toluene conversion. Similar tests were not attempted by BCR, hence no comparison can be made.

Based on an analysis of the products obtained from the simulated recycle tests and on previous experience with similar coals, HRI predict that a steady state continuous process should produce a 43% distillate yield, based on maf coal. An accurate distribution of light distillate, middle distillate and heavy distillate fractions could not be specified. HRI provided a range of probable distributions based on assumptions of the behaviour of the HAO solvent during the hydroliquefaction reaction -- similar to the procedure used by SBRG to define minimum and maximum distillate yields. Depending upon the solvent behaviour, the range of distillate product yields predicted by HRI were:

Light Distillate (IBP - 205°C)	1.8 - 3.7%
Middle Distillate (205°C - 345°C)	5.8 - 19.9%
Heavy Distillate (345°C - 525°C)	35.4 - 19.4%

The detailed HRI report is included in Volume III of this report.

In conclusion, HRI have conducted microautoclave tests with hydrogenated anthracene oil and a once-through recycle solvent (a mixture of HAO and coal-derived solvent). Their results with HAO tend to confirm the BCR microautoclave test results within acceptable experimental accuracy.

Furthermore, the HRI data tend to substantiate the SBRG test results -- that the nominated B.C. coals are difficult to liquefy. The distribution of the liquid fractions in the 43% predicted net distillate yield is based on assumptions related to the behaviour of the anthracene oil in the microautoclave tests. The actual distillate distribution is probably somewhere between the two extremes. It is worth noting that most of the distillate oil is heavy distillate, with a TBP above 345°C. This is very similar to the SBRG predictions.

HRI's predictions of a net distillate yield of 43% for a commercial H-Coal plant are based on results of the microautoclave test program and their experience with similar, but lower rank coals. The highest rank coal that HRI has tested in both microautoclave and continuous bench scale equipment is a high volatile bituminous "A" coal supplied by Ruhrkohle AG, a participant in the H-Coal development program. This German coal, reported to be a typical hard coal with a mean vitrinite reflectance of 0.95-1.0 and a volatile matter content of 34-40%, is lower rank, and therefore a better liquefaction candidate, than the nominated B.C. coals.

It should also be noted that the HRI predictions are based on a vacuum residue recycle operation. In the normal H-Coal mode of operation, the vacuum residue are discharged from the unit, not recycled. Residue recycle operation has never been demonstrated by HRI in large scale continuous pilot plant equipment and, thus, must be considered conceptual, not fully developed.

6.0 PROCESS DEVELOPER PREDICTIONS

6.1 OBJECTIVES

The rationale behind approaching the process developers was to obtain their forecast of the performance of their process using the nominated B.C. coals to produce a synthetic crude product. The quality and the distribution of the products in the primary liquefaction stage were to be important considerations.

Each of the five process developers, HRI, Ruhrkohle/Veba, NCB/BP, Exxon and SRC International, were requested to make their predictions by methods they considered appropriate. These methods could be based on numerical models, batch and/or continuous experiments at any scale with similar coals or their existing analytical data bank. To assist them in assessing their process, the analysis of the two coals, selected test results from the microautoclave tests performed by BCR and, where appropriate, the selected experimental results from the SBRG continuous tests were provided.

The process developers were asked to provide sufficient information to allow development of a scaled plant prediction for conversion of 10 000 tonne per day of coal to liquid products, identifying all product and effluent streams and utility consumptions. In addition to a definition of process operating conditions that would maximize liquid yield of the <190°C fraction, identification of secondary upgrading alternatives which would increase/maximize the liquid product yields was requested.

The developers were requested to comment on the operability of the coals in their raw state and to identify the extent of beneficiation required (e.g. reduction of ash content, increase of volatile matter, increase in reactive/inert content ratio), to maximize liquid yields by their process.

As the Prefeasibility Study progressed and the results of the B.C. Research and SBRG test programs indicated that the nominated B.C. coals were poor liquefaction candidates, follow up discussions were held with three of the process developers, HRI, NCB/BP and Veba Oel (German Process). These meetings led to the establishment of autoclave or extraction test programs with each of the process developers to better evaluate the coals under more optimum process conditions. Results of the HRI test program and NCB predictions are included in this report. The Veba Oel 6 litre autoclave test program will not be completed in time for incorporation into this report.

6.2 H-COAL PROCESS

The detailed information sent to HRI to facilitate their predictions included the coal analysis, non-catalytic and HAO/HDS catalyzed microautoclave test results and details of the SBRG continuous test with Monkman coal and HDS catalyst.

HRI emphasized that the microautoclave tests serve only as a guide to compare the reactivity of different coals at similar conditions and, therefore, the results should not be treated as the basis for a technical evaluation. HRI recognize that the B.C. coals are difficult to liquefy, as was evident from the comparison with Illinois No.6 test results.

During discussions with HRI, it became evident that they did not consider the SBRG test results to be truly representative of the H-Coal process. Even though they appreciated the extent of details of the SBRG procedure, the absence of a recycle coal-derived solvent and the inability to simulate an ebullated bed reactor were of major concern. In HRI's view, a true H-Coal process would be capable of producing higher yields than predicted by the SBRG results. It is recognized that the SBRG tests were not performed at optimized conditions.

By using higher catalyst addition and withdrawal rates (in effect, maintaining a higher catalyst activity), maximum recycle of residual oils and higher reactor temperatures (450 to 460°C versus a normal temperature of 440°C), HRI predicted that 40 to 50% distillate oil yield should be achievable. However, they have no data on similar high rank coals to use as a basis for determining realistic distillate oil yields for the nominated B.C. coals. The highest rank coal processed to date by HRI in both autoclave and continuous bench scale test equipment is a German high volatile bituminous "A" coal.

Based on discussions with HRI, it was decided to commission a microautoclave test program, to be performed by HRI, to better evaluate the H-Coal process with respect to the nominated B.C. coals. Although HRI would have preferred to evaluate the B.C. coals following their standard autoclave and continuous bench scale testing procedures, they agreed to conduct microautoclave tests which HRI claims were designed to improve upon the results of the SBRG test program and to project the estimated yields for the liquefaction of these coals by a representative H-Coal process.

HRI improvised a modification to their microautoclave test program to simulate coal-derived recycle solvent and recycle of vacuum bottoms. With their previous experience on similar coals in the bench and PDU scale tests, HRI were able to predict the yield of a continuous H-Coal process.

Since the initial slurrying solvent used was hydrogenated anthracene oil, a range of yields was predicted based on the behaviour of the anthracene oil during the reaction. The range of the product yields estimated on an maf coal basis are:

Chemcial Gases (H_2O, NH_3, H_2S)	4.9	4.9
Hydrocarbon Gases (C_1 - C_6)	4.9	4.9
Light Distillate (IBP - 205°C)	1.8	3.7
Middle Distillate (205°C - 345°C)	5.8	19.9
Heavy Distillate (345°C - 525°C)	35.4	19.4
Residual Liquid (525 °C +)	37.2	37.2
Unconverted Coal	10.0	10.0
TOTAL	100.0	100.0

Although the distillate yield is expected to be 43%, the distribution of the light, middle and heavy distillate fractions could be anywhere between the two extremes. Even then, most of the distillate oil is in the middle and heavy (205°C +) fractions.

The above predictions are based on HRI's previous experience with 'similar' coals and the concept of vacuum bottoms recycle. As mentioned earlier, the highest rank coal processed by HRI in both autoclave and continuous bench scale test equipment is a German high volatile bituminous "A" coal, which is lower in rank than the B.C. coals. Furthermore, the concept of vacuum bottoms recycle has yet to be demonstrated in continuous pilot scale equipment. Unless the vacuum bottoms recycle concept is successfully demonstrated on a larger scale, this process should be regarded as unproven for liquefying the B.C. coals.

As discussed in Section 5.7, the HRI test results are similar to the B.C. Research microautoclave test results, within acceptable experimental accuracy, and tend to substantiate the SBRG findings that the nominated B.C. coals are difficult to liquefy.

6.3 RUHRKOHLE/VEBA PROCESS

After receiving the test data provided, the German response was somewhat guarded. Although, the B.C. coals could be liquefied by their process, the expected yields would be somewhat low.

To obtain a better understanding of the range of flexibility of the coal properties and get clarification on their guarded statements, discussions were held with Ruhrkohle and the Bergbau-Forschung Research personnel.

The relative low volatile content and high content of non-reactive matter of the B.C. coals were attributed to the prediction of the low distillate yields. This was evident from the BCR microautoclave tests, when the yields of the B.C. coals were compared with Illinois No. 6 coal yields. In addition to the low volatile matter, a significantly important parameter, the B.C. coals possess high indices of the reflectance of the vitrain content (in excess of 1.1). The German process is less suited for coals with reflectance indices in excess of 1.0. The effect of the reflectance indices on the distillate yield is shown in Figure 6.1. The data for Figure 6.1 is based predominantly on German coals and the correlation is valid for more than 12 other coals that have been tested, ranging from lignites, sub-bituminous and bituminous coals from Germany, USA, South America and Australia. The distillate yield for the correlation is based on 4% mineral matter in the dry feed coal.

The effect of mineral content on the yield is shown in Figure 6.2. For the same reactor temperature, the yield reduces substantially with increasing mineral matter content in the coal feed.

Both Figures 6.1 and 6.2 are empirical correlations which are valid for the coals tested by Bergbau-Forschung, but are not necessarily applicable to all coals.

The SBRG continuous tests were performed at 440°C with a solvent on a once through mode of operation. From Figure 6.2, it is clear that higher distillate yields are obtained when the reactor temperature was 474°C than at 464°C. The process developers were asked to comment on the possibility of extrapolation of the curve to 440°C and the confidence thereof. They were unable to make any predictions due to unavailability of data at lower temperatures. Most of their data relate to reactor temperatures at 460°C and higher.

The limitations of the SBRG test results were recognized. Under the conditions available, the SBRG projected yield of between 18% and 34%, represented generally the typical yields that could be obtained from such coals with the use of a proper recycle solvent and higher reactor temperatures. Ruhrkohle speculated that, by reducing the mineral matter content of the coals to about 4%, the net distillate yields would be in the proximity of the maximum yield obtained by the SBRG tests. The speculated yield is based predominantly on the coal proximate analysis. However, definite yield data could be obtained by performing bench scale microautoclave tests followed by continuous tests at their 0.5 tonne/day process development unit at Bergbau-Forschung.

Coals with a variety of properties can be processed in the Ruhrkohle /Veba 200 tonne per day pilot plant, tests on which would be required to obtain the design basis for a commercial plant for any coal. Data for a typical German hard coal are shown in Table 6.1. This coal is the base or reference coal to be processed in the pilot plant. On the same table is shown the flexibility range of coal properties that can be processed in the pilot plant. Although the pilot plant can process a wide range of coals, economic distillate yields would not necessarily be obtained over this entire range.

At a subsequent meeting with Veba Oel, it was decided to commission an autoclave test program to be performed by Veba, to better evaluate the German process with respect to the nominated B.C. coals. The tests are currently underway. However, the results will not be available in time to be incorporated in this report.

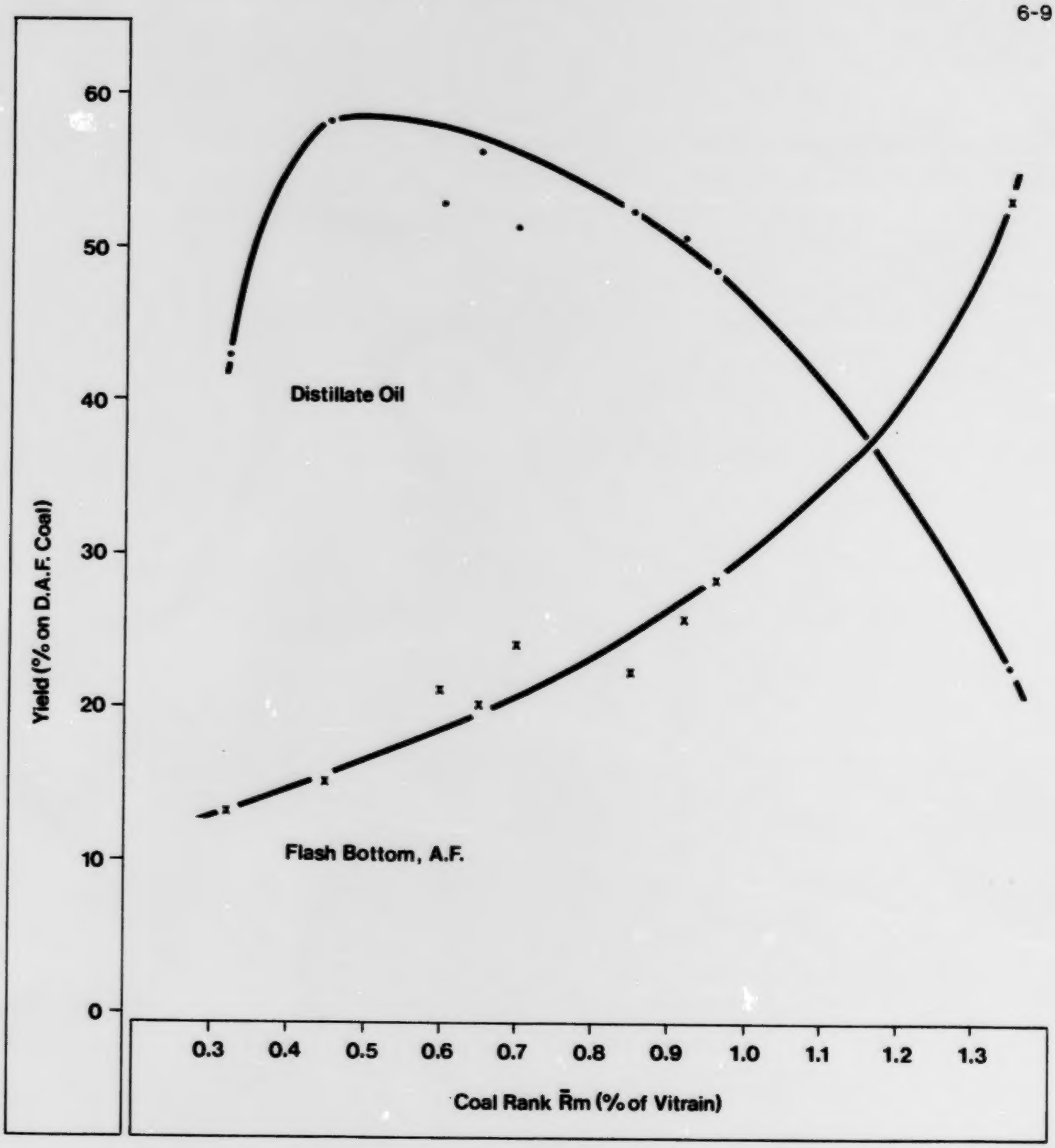


Figure 6.1

Source: "Effects of Some Less Obvious Parameters in Hydroliquefaction"
B. Strobel and F. Friedrich
Bergbau-Forschung

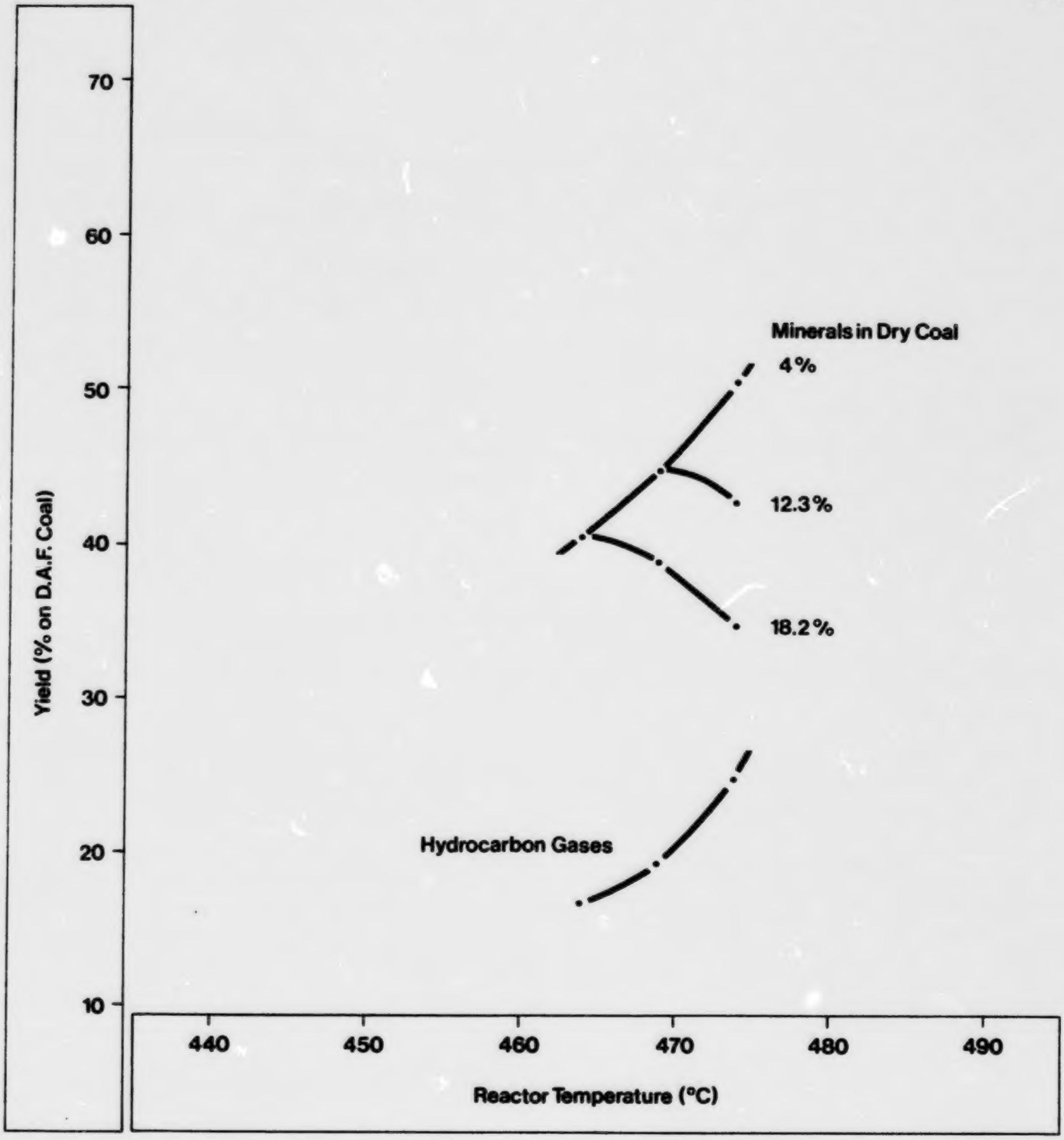


Figure 6.2

Source: "Effects of Some Less Obvious Parameters in Hydroliquefaction"
B. Strobel and F. Friedrich
Bergbau-Forschung

TABLE 6.1

SPECIFICATION OF FEED COALS FOR RUHRKOHLE/VEBA PILOT PLANT

		German Coal	Flexibility
Moisture	wt.%(raw)	4.30	3.8 ... 14.4
Ash	wt.%(m.f.)	4.52	3.0 ... 17.0
Volatile	wt.%(m.a.f.)	37.9	10.0 ... 53.8
Carbon	wt.%(m.a.f.)	84.30	75.0 ... 93.0
Hydrogen	wt.%(m.a.f.)	4.96	3.5 ... 7.0
Oxygen	wt.%(m.a.f.)	7.23	1.5 ... 17.8
Nitrogen	wt.%(m.a.f.)	1.68	1.0 ... 1.9
Sulphur	wt.%(m.a.f.)	1.67	0.5 ... 5.4
Chlorine	wt.%(m.a.f.)	0.16	0.0 ... 0.4
Vitrinite	vol.%	64	61 ... 77
Exinite	vol.%	17	5 ... 20
Inertinite	vol.%	14	1 ... 14
Minerals	vol.%	5	1 ... 9
SiO_2	wt.%	32.92	32.11 ... 64.8
Fe_2O_3	wt.%	32.20	3.71 ... 36.00
Al_2O_3	wt.%	27.63	18.48 ... 34.60
Grindability	°H	52	>30
High Heating Value	kJ/kg(m.a.f.)	34,600	31,500 ... 36,800

Source: "The Coal Hydrogenation Plant at Bottrop " by J. Langhoff, E. Wolowski and G. Escher of Ruhrkohle and Veba Oel.

6.4 NCB/BP LSE PROCESS

As the test programs conducted by B.C. Research and SBRG were not designed to 'simulate' the NCB/BP Liquid Solvent Extraction (LSE) process, discussions were held with the National Coal Board and subsequently with NCB and British Petroleum personnel to obtain information on the expected distillate yields from their process with the nominated B.C. coals and on possible advantages of the process. At the initial meeting between NCB and the Western Canada Coal Liquefaction Study committee, NCB volunteered to conduct extraction tests on samples of the Greenhills and Monkman coals and predict yield data and capital cost for a commerical plant.

Samples of both Greenhills and Monkman coals were sent to the NCB for testing at their research facilities in Stoke Orchard, England. NCB extracted the coals using hydrogenated anthracene oil, and obtained the following yields:

	wt% maf coal		
	Extract	Residue	Gas
Greenhills	72	21	7
Monkman	67	29	4

Using this extraction data and the coal analysis as a data base, NCB were able to make yield and capital cost predictions. NCB claim that these predictions were subsequently confirmed by conducting hydrocracking tests of the extract on their test unit. The predictions are given in Tables 6.2 and 6.3 for an 8 400 tonne per day maf coal feed rate. The extraction and hydrocracking results were not made available to be incorporated in this report.

The National Coal Board appeared to have no doubt that their LSE process could accommodate both Greenhills and Monkman Coal. Indeed, based on the NCB's predictions, this process offers high yields of high quality liquid products.

The yields and economics presented by NCB were for a 50 000 barrel per day facility, approximately 30% larger than that under consideration in this study. Consequently, both yields and capital costs in Tables 6.2 and 6.3 have been pro-rated using Kilborn/Badger in-house data for a facility with the required feed rate of 8 400 tons per day MAF coal. Since the product distribution was provided for only the Greenhills coal an assumption was made for a similar distribution for Monkman coal.

Two cases were presented by NCB, a base case in which coal residue is used for hydrogen generation via gasification, process steam generation and power generation and an alternate case in which hydrogen is produced by steam reforming of natural gas and the coal residue is used for process steam generation and power generation.

NCB assumed that some of the material in the extract would not be easily hydrocracked. Consequently, the configuration they adopted uses the high boiling oil fraction as fuel for both plant and power production. NCB also suggested that further improvement of liquid yields might be possible by coking this fraction where demand existed for high quality coke. However, Exxon's experience with Flexicoking suggests that this may not be a feasible option.

The first quarter 1982 capital cost figures in Tables 6.2 and 6.3 are for a West Coast Canadian location. The figures shown are an initial total cost estimate for all facilities necessary for the total process complex but exclude any mine associated costs. The capital costs provided by NCB included a 10% contingency.

These predictions indicate that the LSE process offers several advantages. The high liquid yields, high quality products, low severity extraction and two stage operation are all attractive features of the process. However, the most limiting feature of this process (at least in comparison to the other processes under consideration) is that to date it has only been demonstrated on a semi-continuous unit with a coal capacity of 30 kg/hr. The next stage of development is to complete the engineering and construction of a 25 tonne per day pilot plant comprising solvent extraction, filtration and hydrocracking units. Once this integrated unit is completed, more data should be available for a better evaluation to be made of the process.

NCB claim that, because of the developed status of the extraction/filtration step and the straight-forward nature of the hydrocracking step, the LSE process is at a stage of development at least as advanced as competing direct liquefaction processes. This claim may be valid on a small scale. However, it should be noted that several of the liquefaction processes, currently being piloted at the 140-200 tonne per day scale, have encountered problems, not necessarily process-derived, that were not envisaged during the laboratory testing stage. Also, NCB/BP have yet to demonstrate an integrated plant on a large scale.

Finally, the success of the NCB LSE process rests on a filtration step which is used to separate mineral matter and unreacted coal from the coal liquid extract. NCB claim that process conditions have been optimized to ensure ease of filtration and this is in no way a difficult or expensive operation. However, this step should be demonstrated on a much larger scale and over a wide range of coals before total confidence can be placed in this process.

TABLE 6.2

PREDICTED NCB YIELDS AND CAPITAL COSTS

Coal Feed Rate 8 400 tonne/day MAF
Base Case - Self Sufficient Plant

	Product Yields	
	Greehills Coal BPD	Monkman Coal BPD
LPG	3 200	3 060
Gasoline	12 260	11 740
Jet Fuel	10 170	9 740
Diesel	9 160	8 770
TOTAL	34 790	33 310
By-product		
Export Power (MW)	-	95
Total Capital Cost	$2320 Million	$2360 Million

Capital cost is in 1982 Canadian Dollars and includes a 10% contingency. This information was prorated and converted to Canadian dollars from data provided by NCB.

TABLE 6.3

PREDICTED NCB YIELDS AND CAPITAL COSTS

Coal Feed Rate 8 400 tonne/day MAF
Alternate Case - Hydrogen by Natural Gas Reforming

	Product Yields	
	Greenhills Coal BPD	Monkman Coal BPD
LPG	3 420	3 060
Gasoline	13 100	11 740
Jet Fuel	10 870	9 740
Diesel	9 790	8 770
TOTAL	37 180	33 310
By-product Export Power (MW)	290	380
Methane Consumption (MMSCFD)	110	88
Total Capital Cost	$1960 Million	$1980 Million

Capital cost is in 1982 Canadian Dollars and includes a 10% contingency. This information was prorated and converted to Canadian dollars from data provided by NCB.

6.5 EDS PROCESS

The initial response of Exxon, after reviewing the coal property data and the non-catalytic tubing bomb test data, was that the B.C. coals could be liquefied. The liquid yields, however, would be lower than for the lower rank coals more commonly processed, e.g. Illinois No. 6, Gulf Coast lignites, Wyoming coal, etc.

Discussions were held with Exxon pertaining to very low conversions for high rank Appalachian Coals which were reported in their 1978-1979 Annual Technical Progress Report. Upon subsequent detailed review, Exxon confirmed that low yields would be expected with high rank (medium and low volatile) coals. As volatile matter content is an important parameter in the EDS process, conversions of the B.C. coals, containing 25.5 and 26.8% volatile matter on a DMMF basis, are expected to be in the range of 20-25% only. Figure 6.3 shows the predicted yields for the B.C. coals, based on the correlation developed by Exxon.

Exxon did stress that the test conditions used to develop the coal data bank and volatile matter versus liquid yield correlation were not optimum for high rank coals. As such, improved liquid yields would be expected if operating conditions were optimized. In addition, Exxon personnel predicted that improved yields would be experienced if residue recycle was incorporated into the process, a step that they are just beginning to develop.

With the very low liquid yields predicted, the EDS process is not recommended for liquefying the B.C. coals.

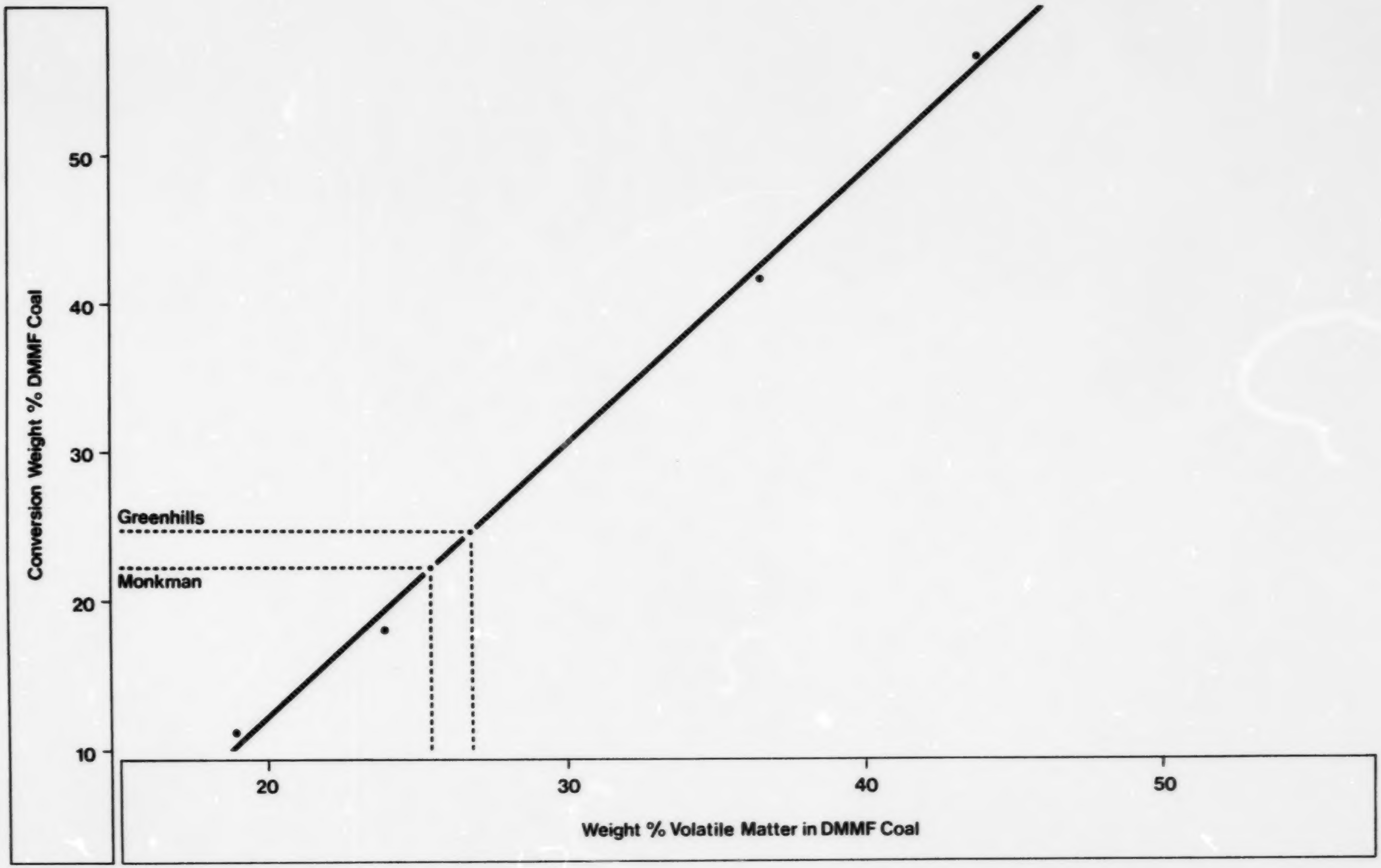


Figure 6.3

Source: Data Extracted from Exxon
Annual Technical Progress Report
1979 - 1980

6.6 SRC II PROCESS

SRC International were contacted for information pertaining to the SRC II process and the expected yields with high rank coals. No response was received -- a further indication that there is no commitment for continued development of the SRC II process since cancellation of the 5 400 tonne per day demonstration plant program late in 1981. In fact the Fort Lewis 45 tonne per day pilot plant is no longer in operation and is reported to be in the process of being dismantled.

Thus, no further evaluation of the SRC II process is recommended.

7.0 PROCESS EVALUATION

7.1 GENERAL

Because of the prediction of low distillate yields (between 19% and 34%) from the SBRG test, the validity of any further evaluation of the coal liquefaction program was questioned. However, in subsequent disucssions with NCB/BP, HRI and Veba Oel, the process developers assured at least a 40% distillate yield. The confidence in the process developers assurance was partially substantiated by the BCR microautoclave test results which predicted distillate yields between 42% and 50% in one set of results (Table 5.1). Furthermore, the SBRG test was performed at one set of predetermined conditions, while the BCR microautoclave tests indicate increased yields with increasing reactor temperature (higher than could be achieved in the SBRG equipment).

Although the microautoclave tests predict the distillate yield, they do not define the split between the naphtha, middle distillate and heavy distillate fractions. To obtain the actual distillate yield and the percentage split of the distillate fractions, further continuous testing would have to be conducted. Time and budget limitations did not permit this extensive testing.

To test the economics of a direct liquefaction process, it was agreed to perform a detailed evaluation with a specific coal and a specific process. The coal selected was from the Greenhills mine since this coal had demonstrated marginally better liquefaction potential in the coal testing program. In no way does this coal selection indicate that possible use of the Monkman coal has been rejected -- further testing must be conducted before a final coal selection can be made. Although any of the three considered processes (H-Coal, Ruhrkohle/Veba or NCB/BP LSE) could have been selected to represent a direct liquefaction plant economics, the H-Coal process was chosen. This decision was made because better in-house information existed on this process. It should be emphasized that the choice of the H-Coal process in no way implies it has any technical and/or economic advantages over the other direct liquefaction processes (Ruhrkohle/Veba or NCB/BP LSE). The selection of this process was made to expedite the prefeasibility study.

A 40% net distillate yield was selected as being representative of the conventional configuration for the H-Coal process feeding Greenhills coal. The evaluation was conducted on the basis of processing Greenhills coal with the exception of the assumed split between naphtha, middle distillate and heavy distillate fractions and the assumed analyses of the raw liquid product fractions. The split between the three liquid fractions was prorated from HRI data on Illinois No. 6 coal to give a net 40% distillate yield and the elemental analyses of the raw products were assumed to be the same as for Illinois No. 6 coal. Consequently, material balances developed in Section 7.5 of this report reflect hydrogen consumption, sulphur production and ammonia production levels for Greenhills coal to synthetic crude. The Base Case design parameters encompass the following:

- An H-Coal liquefaction process operating in the Syncrude mode
- 40% distillate yields
- Hydrogen production by gasification of liquefaction plant residue
- Combined cycle steam/power generation by combustion of liquefaction plant residue
- Rejection of excess liquefaction plant residue in an environmentally acceptable manner.

Development of this Base Case defined the quantities of intermediates and products. This enabled the various alternative processing configurations to be selected, any of which could possibly offer improved economics. The alternative processing schemes were then developed to a stage sufficient to produce preliminary comparative economics for each. The alternative schemes investigated were:

Case 1) Base Case conditions, but with 50% distillate yields.

Case 2) Base Case conditions, but with gasification of excess residue and partial oxidation of excess fuel gas to produce a synthesis gas to be converted to gasoline via the Mobil MTG process. In this configuration electricity would be purchased.

Case 3) As in Case 2, but with gasification of the coal beneficiation plant rejects in addition to excess residue to enhance synthesis gas production and hence increase gasoline production via the MTG route.

Case 4) A 40% distillate yield with hydrogen production via steam reforming of purchased natural gas and any available excess process fuel gas. In this case the liquefaction plant residue and coal beneficiation plant rejects are gasified and synthesized to produce gasoline via the MTG process.

Case 5) A 40% distillate yield with hydrogen production via steam reforming of natural gas and any available excess process fuel gas. In this case, the liquefaction plant residue will be used for plant steam and power generation. Excess residue will be disposed of at an assumed zero disposal cost.

The above alternatives were investigated from an economic standpoint only. The economics for the Base Case and alternative configurations are presented in Volume II of this report. The detailed description of the Base Case configuration is presented in Sections

7.2 to 7.6. Section 7.7 gives a brief description and simplified block flow diagram for each of the alternatives investigated. A tabulated comparison of the different process configurations is given in Section 7.7.

7.2 PROCESS EVALUATION PARAMETERS H-COAL - BASE CASE

7.2.1 Synthetic Crude Production

7.2.1.1 Naphtha Fraction

Production, tonne year	338 830
Production, BBLS/day	9 045
Specific Gravity	0.76
Boiling Range, °C	C_5 - 200
Sulphur, ppm	1
Nitrogen, ppm	1
Oxygen, ppm	30

7.2.1.2 Mid Distillate Fraction

Production, tonne/year	483 430
Production, BBLS/day	11 020
Specific Gravity	0.89
Boiling Range, °C	200-350
Sulphur, ppm	50
Nitrogen, ppm	1 000
Oxygen, ppm	1 500

7.2.1.3 Heavy Distillate Fraction

Production, tonne/year	199 430
Production, BBLS/day	4 215
Specific Gravity	0.96
Boiling Range, °C	350-525
Sulphur, ppm	100
Nitrogen, ppm	2 000
Oxygen, ppm	2 000

7.2.1.4 Butane

Production, tonne/year	83 960
Production, BBLS/day	2 940
Specific Gravity	0.58
Propane, LV% max.	2.0
Butanes, LV% min.	97.0
C_5's, LV% max.	2.0
Sulphur, Doctor Test	Sweet

7.2.1.5 Propane

Production, tonne/year	92 370
Production, BBLS/day	3 725
Specific Gravity	0.50
Ethane, LV% max.	1.5
Propane, LV% min.	98.0
Butanes, LV% max.	1.5
Sulphur, Doctor Test	Sweet

7.2.2 Area 200 - Coal Feedstock

7.2.2.1 Coal "As Received"

Feed Rate, tonne/hr	417
Feed Rate, tonne/year	3 100 000
Moisture Content, wt%	8.0
Ash Content, wt% (as received basis)	8.0
Ash Content, wt% (dry basis)	8.7

7.2.3 Area 300 - Coal Liquefaction

7.2.3.1 Coal Crushing and Drying

Feed Rate, tonne/hr design	425
Feed Coal Particle Size, mm.	minus 50
Feed Coal Moisture, wt%	8
Ground Coal Particle Size, mesh	minus 20
Dried Coal Moisture, wt%	2

7.2.3.2 Coal Slurry Preparation

Coal Feed Rate, tonne/hr (8% ash, 2% moisture)	391
Coal Size Range, mesh	minus 20
Delivery Method	pneumatic

Coal Analysis , wt% Dry Basis

Carbon	78.88
Hydrogen	4.42
Nitrogen	1.05
Sulphur	0.40
Oxygen	6.55
Ash	8.70
Moisture Content wt%	2
Slurry Oil Recycle to slurry prep vessel, tonne/hr	390
Slurry Oil Specific Gravity	1.038
Slurry Oil Temperature °C	300
Solids Content of Slurry wt%	50

7.2.3.3 Reaction Section

Coal Slurry Oil to Reactor, tonne/hr	1 222
Solids Content of Coal Slurry to Reactor, wt%	30
Reactor Inlet Temperature	*
Reactor Inlet Pressure	*
Reactor Length (T-T), m	17.7
Reactor Design Temperature, °C	470
Reactor Design Pressure, kPa	22 100
Reactor Wall Construction	2 1/4Cr-1 Mo 347SS Lining 150 mm Refractory
Reactors Per Train	1
Total Number of Trains	5
Hydrogen Purity, mol% min.	95
Total Hydrogen to Reactors (as 100% H_2) Nm³/kg maf coal	0.6
Liquid Residue Let-Down System	2 Stage
First Stage, kPa	20 500-8 000
Second Stage, kPa	8 000-500

Product Yields

	wt% maf Coal	wt% mf Coal
H_2S	0.33	0.30
NH_3	1.02	0.93
CO	0.18	0.16
CO_2	0.40	0.37
H_2O	5.39	4.92
C_1	4.05	3.70
C_2	3.88	3.54
C_3	3.71	3.39
C_4	2.67	2.43
C_5 - 200°C	12.00	10.96

* Proprietary Information

Product Yields (Cont'd)

	wt% maf Coal	wt% mf Coal
200°C - 350°C	17.64	16.10
350°C - 525°C	13.54	12.37
Residue HC	34.22	31.25
Unreacted Coal	6.33	5.78
Ash		8.69
	105.36	104.89
Hydrogen Consumed	5.36	4.89
	100.00	100.00

7.2.4 Area 400 Atmospheric and Vacuum Distillation

7.2.4.1 Fractionation

Total Feed to Fractionation, tonne/hr	822.7
Stabilized Heavy Naphtha	
Flow Rate, tonne/hr	31.5
Flow Rate, BBLS/day	6 176
Specific Gravity	0.77
Boiling Range, °C	C_5 -200
Ultimate Analysis, wt%	
C	85.34
H	13.79
N	0.24
S	0.07
O	0.55
Raw Middle Distillate	
Flow Rate, tonne/hr	60.0
Flow Rate, BBLS/day	9 600
Specific Gravity	0.943
Boiling Range, °C	200-350
Ultimate Analysis, wt%	
C	88.35
H	10.11
N	0.47
S	0.08
O	0.99

7.2.4.1 Fractionation (Cont'd)

Raw Heavy Distillate	
Total Flow Rate (incl. recycle oil), tonne/hr	732.1
Total Flow Rate (incl. recycle oil), BBLS/day	106 470
Net Heavy Distillate, tonne/hr	38.7
Net Heavy Distillate, BBLS/day	5 627
Specific Gravity	1.038
Boiling Range	350-525
Ultimate Analysis, wt%	
C	89.42
H	8.56
N	0.62
S	0.08
O	1.28

7.2.4.2 Solids Separation

Total Feed to Steam Stripper, tonne/hr	380.7
Residue	
Total Flow Rate, tonne/hr	186.4
Residue Analysis, wt%	
Solid Hydrocarbon Residue	64.3
Unreacted Coal	11.9
Heavy Distillate	6.0
Ash	17.9
Ultimate Analysis, wt%	
C	91.8
H	4.6
N	0.1
S	0.2
O	3.3

7.2.5 Area 500 - Naphtha Hydrotreater

7.2.5.1 Raw Naphtha

Feed Rate, tonne/hr	46.2
Feed Rate, BBLS/day	9 030
Specific Gravity	0.77
Boiling Range	C_5 - 200
Hydrogen Addition to Raw Naphtha (as 100% H_2), Nm³/m³	76
Hydrogen Addition to Raw Naphtha (as 100% H_2), SCF/BBL	428
Estimated Yields wt%	
H_2O	0.62
NH_3	0.29
H_2S	0.07
C_1 - C_4	2.89
C_5 - 200	97.02
	100.89

7.2.5.2 Treated Naphtha

Flow Rate, tonne/hr	45.5
Flow Rate, BBLS/day	9 045
Specific Gravity	0.76
Boiling Range, °C	C_5 - 200
Oxygen, ppm	30
Nitrogen, ppm	1
Sulphur, ppm	1

7.2.6 Area 600 - Middle Distillate Hydrotreater

7.2.6.1 Raw Middle Distillate

Feed Rate, tonne/hr	66.3
Feed Rate, BBLS/day	10 600
Specific Gravity	0.94
Boiling Range, °C	200-350

7.2.6.1 Raw Middle Distillate (Cont'd)

Hydrogen Addition to Raw Middle Distillate (as 100%), Nm^3/m^3	200
Hydrogen Addition to Raw Middle Distillate (as 100%), SCF/BBL	1 114
Estimated Yields (wt%)	
H_2O	0.95
NH_3	0.68
H_2S	0.08
C_1	0.30
C_2	0.48
C_3	0.48
C_4	0.96
C_5-200	2.00
200-350	96.01
	101.94

7.2.6.2 Treated Middle Distillate

Flow Rate, tonne/hr	65.0
Flow Rate, BBLS/day	11 020
Specific Gravity	0.89
Boiling Range, °C	200-350
Oxygen, ppm	1 500
Nitrogen, ppm	1 000
Sulphur, ppm	50

7.2.7 Area 700 - Heavy Distillate Hydrotreater

7.2.7.1 Raw Heavy Distillate

Feed Rate, tonne/hr	38.7
Feed Rate, BBLS/day	5 640
Specific Gravity	1.038
Boiling Range, °C	350-525

7.2.7.1 Raw Heavy Distillate (Cont'd)

Hydrogen Addition to Raw Heavy Distillate (as 100%), Nm³/m³	869
Hydrogen Addition to Raw Heavy Distillate (as 100%), SCF/BBL	4 792
Estimated Yields (wt%)	
H_2O	1.20
NH_3	0.75
H_2S	0.07
C_1	1.62
C_2	1.91
C_3	1.65
C_4	3.30
C_5 - 200	12.00
200- 350	19.00
350- 525	66.00
	107.50

7.2.7.2 Treated Heavy Distillate

Flow Rate, tonne/hr	26.8
Flow Rate, BBLS/day	4 215
Specific Gravity	0.96
Boiling Range, °C	350-525
Oxygen, ppm	2 000
Nitrogen, ppm	2 000
Sulphur, ppm	100

7.2.8 Area 800 - Light Ends Recovery

7.2.8.1 Sweet Gas Feed

Flow Rate, tonne/hr	79.6
Flow Rate, Nm³/hr	92 800
Pressure, kPa	2 450
Temperature, °C	45

7.2.8.2 Product Streams

High Purity Hydrogen

Flow Rate, tonne/hr	2.0
Flow Rate, Nm^3/hr	22 000
Hydrogen Content, wt%	99.5
Temperature, °C	15
Pressure, kPa	2 100

Fuel Gas

Flow Rate, tonne/hr	42.3
Flow Rate, Nm^3/hr	46 800
Heating Value, KJ/Nm^3 (LHV)	44 035
Heating Value, BTU/SCF (LHV)	1 120
Molecular Weight	20.7
Temperature, °C	10
Pressure, kPa	375

Propane

Flow Rate, tonne/hr	12.4
Flow Rate, BBLS/day	3 725
Specific Gravity	0.50
Ethane, LV% max.	1.5
Propane, LV% min.	98.0
Butanes, LV% max.	1.5
Sulphur, Doctor Test	Sweet

Butane

Flow Rate, tonne/hr	11.3
Flow Rate, BBLS/day	2 940
Specific Gravity	0.58
Propane, LV% max.	2.0
Butanes, LV% min.	97.0
C_5's, LV% max.	2.0
Sulphur, Doctor Test	Sweet

Stabilized Light Naphtha

Flow Rate, tonne/hr	10.2
Flow Rate, BBLS/day	2 260
Specific Gravity	0.68
Boiling Range	C_5 -200

7.2.9 Area 900 - Sulphur Plant

7.2.9.1 Sulphur Plant Feed

Feed Rate, tonne/hr	38.7
Inlet Temperature, °C	50
Inlet Pressure, kPa	515
Feed Composition, mol%	
C_1	18.51
C_2	8.48
C_3	7.27
C_4	8.33
C_5^+	5.62
H_2	17.08
H_2O	9.35
CO	9.34
CO_2	12.66
H_2S	3.66
	100.00
Sulphur Recovery, %min	98

7.2.9.2 Sulphur Product

Flow Rate, tonne/hr	1.44
Sulphur Content, wt% min.	99.8
H_2S Content, ppm	< 1
Color	BRIGHT YELLOW
H_2S Content in Sulphur Plant Off Gas	<10 ppm

7.2.10 Area 1 000 - Hydrogen Production

7.2.10.1 Gasification Section (Texaco Gasifier)

Residue Feed Rate, tonne/hr	132.3
Oxygen Feed Rate, tonne/hr	81.3
Temperature, °C	1 400
Pressure, kPa	4 000
Wet Ash Reject Rate, tonne/hr	31.6
Ash Moisture Content, wt%	25

Quenched Gas Composition

	wt %	mol %
Methane	1.54	2.56
Hydrogen	2.65	35.27
H_2O	35.03	51.85
CO	55.78	7.14
CO_2	4.50	2.72
H_2S	0.07	0.06
N_2 + Ar	0.43	0.40

7.2.10.2 CO Shift Conversion Section

Wet Gas Feed Rate, tonne/hr	372.1
Temperature - mean, °C	350
Pressure, kPa	3 750

Shifted Gas Composition

	wt %	mol %
Methane	1.54	1.79
Hydrogen	6.29	58.68
CO	1.26	0.84
CO_2	90.41	38.37
N_2/Ar	0.07	0.04
H_2S	0.43	0.28

7.2.10.3 Sour Gas Treatment

Gas Feed Rate, tonne/hr	372.1
Hydrogen Production	
Flow Rate, tonne/hr	37.3
Flow Rate (as 100% H_2), tonne/hr	23.4

Treated Gas Composition

	wt %	mol %
Methane	14.81	2.81
Hydrogen	62.66	95.28
CO	8.95	0.97
CO_2	13.58	0.94
	100.	100.

7.2.11 Area 1 100 Oxygen Plant

Oxygen Plant Capacity (as 100% O_2) design, tonne/hr	87
Oxygen Purity, min. wt%	99.5
Nitrogen Purity, wt%	99.99

7.2.12 Area 1 300 and Area 1 400 Phenol and Ammonia Recovery

Flow Rate, tonne/hr	300
Composition, wt%	
CO_2	0.14
NH_3	1.61
H_2S	0.08
Phenols	0.13
Water	98.04
Phenolic Removal, % min.	99.0
Anhydrous Ammonia Product, tonne/hr	4.8

By-product Utilization:

The separated Phenolics are burned as fuel in a fired heater while Sodium Carbonate is sold as a by-product.

7.3 H-COAL PROCESS DESCRIPTION - BASE CASE CONFIGURATION

7.3.1 General

Beneficiated ground coal at a rate of 8 400 tonnes per day on a maf basis is fed to the liquefaction unit. The liquids produced by the liquefaction unit are separated into light, middle and heavy fractions and upgraded by hydrotreating to produce syncrude product. In addition, propane and butanes are produced. The lighter gases (C_2-) produced are utilized within the complex as fuel gas.

Make-up hydrogen for the liquefaction unit and product upgrading is produced by gasification of the liquefaction plant residue. The oxygen for the gasifiers is also produced within the complex. The remaining liquefaction plant residue is granulated and part of it is fed to boilers for steam and power generation. Any excess residue is disposed of in an environmentally acceptable manner. Incineration is considered to be a possible method of disposal.

The Base Case configuration comprises the following units:

Coal Preparation
Coal Liquefaction
Product Separation
Product Upgrading
Light Ends Recovery
Hydrogen Production
Oxygen Production
Effluent and Emmission Control
Offsites

The H-Coal Block Flow Diagram 100-43-001, presented in Section 7.4 of this report, illustrates the inter-relationship between the process units within the complex.

7.3.2 Area 300 - Coal Liquefaction

Conceptual Flow Diagram 100-43-002 illustrates the major equipment in a single train of the liquefaction section of the plant.

7.3.2.1 Coal Preparation

Beneficiated feed coal is discharged onto a conveyor from the storage silo by variable speed feeders which control the coal flow. The coal is weighed and discharged into the pulverizer feed hopper which has a thirty minute hold-up time. Variable capacity feeders in the hopper discharge the coal into two bowl mill pulverizers. As the coal is pulverized, it flows to the outer edge of the bowl and is picked up by a stream of hot combustion gas and conveyed to a classifier inside the mill housing where smaller sizes are conveyed to cyclone separators and the oversize particles are returned to the bowl. Bag-house filters are provided to handle the net gas streams from the cyclones and to limit particulate emissions. Process feed coal is classed as 20 mesh. From the cyclone separators, the coal flows to two pulverized coal receivers from which it is fed by screw feeder to the liquefaction feed slurry tanks.

7.3.2.2 Coal Liquefaction

The coal liquefaction unit consists of five parallel and identical modules for coal slurrying, liquefaction, high pressure vapour cooling and intermediate and low pressure cooling and flashing. Each module is designed to handle one-fifth of the coal feed. Vapour streams from the five modules are recombined as feed to the high pressure acid gas removal and hydrogen recovery system.

The ground coal is blended with the heavy distillate recycle oil in the slurry preparation vessel to a 50 wt% coal slurry. The coal slurry is suspended by agitation within the preparation vessel and recycling slurry. The temperature of the vessel contents is kept at 170°C (340°F) to allow for rapid dispersion of the coal in the oil and maximum heat recovery.

The 50% slurry is pumped from the slurry preparation vessel to a slurry holding vessel maintained at 250°C and 2 600 kPa where the remaining heavy distillate recycle oil and the lean solids overflow from the hydroclones is recycled to maximize coal conversion and control viscosity of coal slurry as it is being heated. Slurry from the holding tank, approximately 30 wt% solids, is pumped to the suction of the high pressure booster pumps by centrifugal pumps. The slurry is pressurized to a reactor pressure of 21 500 kPa (3 150 psig) and mixed with about one-half the required quantity of recycle and make-up hydrogen.

During heating of the coal slurry in the fired preheater the slurry viscosity increases and passes through a maximum at a temperature of about 350-400°C. About one-half of the hydrogen stream is added to the coal slurry prior to entering the heater to minimize laminar flow conditions caused by the high viscosity of the slurry and to prevent coking of the heater tubes.

The remaining hydrogen stream is heated in a separate fired heater to reaction conditions and introduced into the bottom of the liquefaction reactor. The coal slurry/hydrogen stream is introduced into the bottom of the reactor separately.

The H-Coal reactor liquefies coal at approximately 20 500 kPa (3 000 psig) and 455°C (850°F) using the principle of an ebullated bed of catalyst. Ebullation is achieved by the upflowing coal/oil slurry and gas streams and internal liquid recycle which expands the catalyst bed and fluidizes the particles. Due to bed expansion the reactor pressure drop is small and rapid mixing within the reactor is achieved.

The ebullated bed reactor permits higher feed rates at lower pressure drops than is possible with conventional fixed bed catalyst reactors and provides for intimate contacting of the reactants and catalyst. The heat of reaction is dissipated rapidly within the reactor allowing for near isothermal temperature conditions and close control of the operating temperature. Rapid backmixing also utilizes the heat of reaction to raise the temperature of the reactants to the reactor temperature.

A common problem in fixed bed catalytic hydrogenation processes is coking of the reactor walls and catalyst particles. Rapid mixing within the ebullated bed reactor allows catalyst particles to rub against each other and the walls of the reactor, thus minimizing the formation of deposits on the catalyst surface and reactor wall. This can be of particular importance when dealing with coals which have a fouling tendency or contain high quantities of certain minerals, such as calcium, which are prone to deposit.

Unreacted coal and ash particles, being lighter than the catalyst particles, are carried out of the reactor in the gas-liquid effluent. The catalyst bed expansion is controlled to keep the catalyst within the reactor. With time the catalyst becomes deactivated and, being heavier, tends to settle to the bottom of the reactor. To maintain catalyst activity within the reactor deactivated catalyst is removed periodically from the reactor and replaced with fresh catalyst. The catalyst used is a cobalt-molybdenum hydrodesulphurization catalyst.

Catalyst addition and removal is achieved on-line using a lockhopper type of operation. Fresh catalyst is added to a catalyst hopper and drained by gravity into a catalyst addition drum. The drum is pressurized with make-up hydrogen to a pressure slightly above the reactor pressure. The valve to the reactor is opened and catalyst is transferred into the reactor catalyst bed through a stand pipe in the reactor top head.

Spent catalyst is withdrawn from the bottom of the catalyst bed using a catalyst removal drum. The removal drum is pressurized to reactor pressure and preheated to avoid thermal shock. On average about 16 tonne/day of catalyst must be withdrawn in this manner and replaced with fresh catalyst.

7.3.2.3 Reactor Effluent Separation

The combined gaseous, liquid and solid reactor effluent is received in the hot effluent separator where an initial separation of vapour and gases from liquid and solids is achieved. Hot gases and vapour from the hot separator are cooled to 260°C by preheating the hydrogen recycle gas stream and the fractionator feed stream. Condensate produced by this cooling is separated in a knockout drum and the remaining vapour, still at high pressure (20 000 kPa), is cooled to 70°C by producing steam, heating boiler feed water and air cooling.

Wash water is injected into the vapour at the inlet to the boiler feed water preheater to ensure that ammonium salts are dissolved and do not plug downstream exchangers or lines. The cooled effluent is separated into vapour, light hydrocarbons and an aqueous phase.

This two-stage condensation of vapour is used throughout the product separation area wherever addition of wash water is required to remove ammonium salts. The first condensation train cools the vapour to a relatively high temperature to condense and separate the heavy hydrocarbons in the vapour stream before injection of the wash water. This is done to avoid the creation of an emulsion of heavy hydrocarbons in the aqueous phase. The cooled vapour stream is compressed and recycled to the reactors.

Reactor effluent slurry from the hot separator is flashed to 8 000 kPa to recover dissolved hydrogen and thereby minimize the quantity of hydrogen make-up required. The pressure of the flash is set so as to minimize hydrogen recompression and to allow for a two-stage reduction in slurry pressure which tends to minimize erosion wear in the slurry let-down valves. The flashed slurry is quenched to 425°C by direct contact with subcooled hydrocarbon condensate to avoid the possibility of coke formation.

Vapour from the 8 000 kPa slurry flash is combined with flashed, condensed hydrocarbon at 260°C from the first stage of the vapour cooling, cooled to 220°C by generating steam and fed to a condensate drum. A portion of the condensate drum liquid is used to quench hot flashed slurry. Vapours from the condensate drum at 8 000 kPa and 220°C are quenched with water to avoid ammonium salts precipitation and cooled to 50°C by air-cooling.

The cooled effluent from the air-cooler is mixed with flashed hydrocarbon condensate from the second stage of vapour cooling and fed to a condensate wash water separator. Vapour, hydrocarbon liquid and an aqueous phase are separated to 8 000 kPa. The sour aqueous phase is separated, flashed to 350 kPa and sent to a water degassing drum and eventually to the ammonia recovery unit. Vapours are sent to the high pressure acid gas treatment system.

Slurry from the 8 000 kPa flash drum is further flashed to a 600 kPa flash drum. Slurry in this drum is kept in suspension by external pump recirculation. Recovered vapour is cooled by preheating the fractionator feed stream.

Vapour from the low pressure condensate drum at 275°C is cooled by addition of wash water to dissolve ammonium salts present and then cooled by air-cooler to 50°C. The air-cooler effluent is mixed with recovered hydrocarbon from the 8 000 kPa condensate was water separator and fed to a low pressure (500 kPa) condensate wash water separator drum. The aqueous phase containing ammonium salts is fed to the water degassing drum and eventually to the ammonia recovery unit. Vapours from the low pressure separator are sent to the sour gas treatment unit. Recovered hydrocarbon is mixed with hydrocarbon from the low pressure condensate drum to form part of the fractionator feed.

Slurry in the 660 kPa flash drum contains all the solids remaining in the reactor effluent after liquefaction. This slurry, at 410°C, is sent to a hydroclone system to reduce the concentration of solids in that portion of the recycle oil slurry required to control the composition of the heavy ends in the reactor feed slurry.

The hydroclone overflow stream, as a low solids stream, is flashed to 200 kPa to reduce its vapour pressure prior to recycling to the slurry preparation area. Total recycle slurry oil is made up of approximately 50% heavy distillate and 50% hydroclone overflow. Hydroclone underflow, as the rich solids stream, is pumped to the solids separation unit.

7.3.2.4 High Pressure Sour Gas Treatment and Hydrogen Recovery

Gases from 8 000 kPa wash water separator is fed to the high pressure sour gas treatment and hydrogen recovery unit. The purpose of this unit is to provide an enriched hydrogen stream for recycle to the coal conversion unit.

Vapour and gases at 8 000 kPa are first sent to an amine unit to remove acid gases, carbon dioxide and hydrogen sulphide. This unit uses diglycolamine to combine with acid gases in the amine absorber. Spent amine is withdrawn from the bottom of the absorber and sent to an amine stripper where the gases are released and transferred to the sulphur recovery plant. Regenerated amine is recycled to the absorber after cooling.

Sweet gases from the amine absorber are passed through a molecular sieve bed to remove trace contaminants and fed to a cryogenic hydrogen recovery unit. In the cryogenic unit hydrogen is recovered from the gas stream. Enriched hydrogen, at 92% purity from the cryogenic unit is distributed to each of the five modules, where it is compressed and recombined with the hydrogen stream from the 20 000 kPa wash water separator. This recombined recycle hydrogen stream is mixed with the make-up hydrogen stream to form the hydrogen feed to the coal liquefaction unit. Waste gases from the cryogenic unit are transferred to the light ends recovery unit.

7.3.3 Area 400 - Product Separation

Conceptual Flow Diagram 100-43-003 illustrates the major equipment in the production separation section of the plant.

7.3.3.1 Distillation

Fractionator feed from each module is preheated successively by hot reactor effluent and low pressure flash vapour to 310°C. Final feed preheating of the combined stream takes place in a fractionator fired heater in which the feed is heated to 340°C. Fractionator flash zone conditions are typically set at 320°C (610°F) and 375 kPa (55 psia) to permit maximum recovery of bottoms product. Stripping steam, at a rate of about 30 kg per cubic metre of bottoms product (11 lb per barrel of bottoms product), is introduced into the bottom of the fractionator to lower the flash point of the bottoms product and reduce the hydrocarbon partial pressure in the flash zone.

Inert gases, light hydrocarbons and naphtha (C_5-200°C) are taken overhead in the fractionator, cooled by air cooling and trim water coolers to 40°C and separated in an overhead accumulator. Off gases from the overhead accumulator are transferred to the sour gas unit for acid gas removal. Light naphtha liquid product is removed from the accumulator and routed to the naphtha stabilizer. Sour water from steam stripping is transferred from the accumulator to the ammonia recovery unit.

Two middle distillate sidecuts are withdrawn from the fractionator and passed through two sidecut strippers for light ends removal by steam stripping. Middle distillate is cooled by air-cooling and transferred to the middle distillate hydrotreater for upgrading at a rate of 60.0 tonne/hr.

A portion of the fractionator bottoms stream is combined with the distillate from the vacuum tower (Section 7.3.3.3), air cooled and transferred to the heavy distillate hydrotreater for upgrading. The remainder of the stream is distributed to each of the five coal conversion units, where it is blended with the hydroclone overflow to form the recycle oil for coal slurrying.

7.3.3.2 Naphtha Stabilizer

Naphtha stabilizer feed is preheated by the stabilizer bottoms from 60°C to 120°C and fed to the naphtha stabilizer. Stabilized naphtha from the bottoms is used to preheat the feed and is then fed to the naphtha hydrotreater at a rate of 31.5 tonne/hr. Light gases from the stabilizer form a part of the sour gas to the sour gas treatment unit.

7.3.3.3 Solids Separation

Combined low pressure hot slurry from the hydroclone underflow of the five modules is stripped of low boiling hydrocarbons in a steam stripper. Overhead vapour from the stripper is mixed with vapour from the combined hydroclone overflow flash drums and cooled by preheating the boiler feed water and trim air cooling to 90°C. The partially condensed overhead stream is fed to a hydrocarbon/water separator. Sour water from the accumulator is transferred to the ammonia recovery unit. Hydrocarbon from the steam stripper overhead separator is distributed to each coal conversion unit where it is combined with hydrocarbon from the low pressure wash water separators and low pressure flash drums and preheated by cooling the hot reactor vapour effluent. These streams are then recombined to form the fractionator feed stream.

The slurry recovered as bottoms from the steam stripper is fed to the vacuum tower. By vacuum distillation at 5 kPa, the heavy distillate is recovered for blending with fractionator heavy distillate.

7.3.3.4 Residue Granulation

The residue required for hydrogen production is pumped directly to the gasifiers. The excess residue from the bottom of the vacuum tower is dropped by gravity to a residue slating system where it is cooled from 375°C to 70°C. Solidification of the residue occurs at about 200°C. In the residue slating system, residue is spread in a thin layer over a moving stainless steel belt. A water spray system beneath the belt is used to cool the belt and the residue. The solidified residue passes through a breaking device as it leaves the belt and is broken into minus 50 mm (minus 2 in.) pieces. Granulated residue is transferred on a belt conveyor to the steam and power generation plant and/or to disposal as required.

7.3.4 Areas 500, 600 and 700 - Product Upgrading

All liquid products from the coal liquefaction process are hydrotreated for stability. This is done by reducing the sulphur, oxygen and nitrogen content of the liquids using UOP hydrotreating processes. Details on the status of development of liquid fraction upgrading options are given in Section 4.7 of this report.

7.3.4.1 Area 500 - Naphtha Hydrotreating

Naphtha, at a rate of 31.5 tonne/hr from the naphtha stabilizer, 10.2 tonne/hr of light naphtha from light ends recovery and 4.5 tonne/hr of hydrocracked naphtha from the heavy distillate hydrotreater, are processed in the naphtha hydrotreater to yield a stable naphtha product suitable for refinery use as a catalytic reformer feedstock.

The UOP LT Unibon Process was selected as UOP Process Division has considerable commercial experience in treating cracked naphthas (FCC, Thermal Cracker, Coker, Visbreaker) and pyrolysis naphthas and has conducted a series of pilot plant studies on hydrotreating coal-derived naphthas (H-Coal, Exxon Donor Solvent, SRC-II).

The UOP LT Unibon Process is a two-stage hydrotreating process employing suitable catalysts to allow the selective hydrogenation of a wide range of diolefins, monoolefins, sulphur, nitrogen and oxygen contaminants without the formation of troublesome polymer, gum and coke on heat exchangers, heater tubes and catalyst beds. Hydrocracking of feed to light ends is minimal and is confined to those fragments created by the fracturing of the sulphur, nitrogen and oxygen containing compounds.

In the LT Unibon Unit, naphtha feedstock is combined with recycle and makeup hydrogen, preheated via process heat exchange with reactor effluent and charged to the prereactor. The catalyst used in this stage is extremely active and selective and achieves the rapid hydrogenation of dienes, styrene type materials and other unsaturated compounds in the feed to the corresponding paraffins, aromatics or naphthenes-even at low temperature. Polymerization and condensation reactions are suppressed, thereby minimizing the fouling of exchangers and catalyst beds.

The prereactor effluent stream is heated in a fired heater and charged to the main reactor to effect the removal of sulphur, nitrogen and oxygen compounds. The catalyst and operating conditions for this reactor have been selected to combine maximum removal of sulphur, nitrogen and oxygen contaminants with maximum retention of aromatic and naphthenic compounds.

Hydrocracked naphtha produced in the Heavy Distillate Hydrotreater has sulphur, nitrogen and oxygen contents moderately in excess of desired product limits. This naphtha stream is heated in the convection section of the main fired heater and charged to the main reactor at a point part way through the catalyst bed. Only mild hydrotreating is required for this naphtha stream to meet product specifications.

The effluent from the main reactor is cooled and fed to a high pressure separator. The hydrogen-rich separator gas is recycled to the reactors and makeup hydrogen is added to this gas stream to maintain hydrogen partial pressure in the system as hydrogen is consumed in the process. The separator liquid is charged to a naphtha stabilizer where H_2S, NH_3 and water formed in the process are removed as an overhead cut together with a small quantity of C_4-hydrocarbons. The stabilizer bottoms product is cooled and pumped to product storage at a rate of 45.5 tonne/hr (9 045 barrels per day).

7.3.4.2 Area 600 - Middle Distillate Hydrotreater

Middle distillate, 200-350°C, from the atmospheric fractionator at a rate of 60.0 tonne/hr and 6.3 tonne/hr of hydrocracked middle distillate from the Heavy Distillate Hydrotreater are processed in the Middle Distillate Hydrotreater to produce a treated middle distillate product, low in sulphur, nitrogen and oxygen, which is suitable as Fluid Catalytic Cracker feedstock or as industrial boiler or turbine fuel or for further upgrading. Critical product specifications include a maximum oxygen concentration of 1 500 ppm, a maximum nitrogen concentration of 1 000 ppm and a sulphur content <50 ppm.

The UOP HD Unibon Process was selected as UOP Process Division has considerable commercial experience in treating middle distillate refinery stocks for sulphur, nitrogen and oxygen removal and has conducted a series of pilot plant studies on hydrotreating, Fluid Catalytic Cracking and hydrocracking coal-derived gas oils (H-Coal, Exxon Donor Solvent, SRC-II).

The UOP HD Unibon Process is a single-stage hydrotreating process employing suitable catalysts to allow selective reaction of sulphur, nitrogen and oxygen containing compounds with hydrogen. The removal of contaminants involves the controlled breaking of the molecular chain or ring at the point where the sulphur, nitrogen or oxygen atom is joined to carbon atoms. This breaking is accomplished by the introduction of hydrogen with production of hydrogen sulphide, ammonia and water. The resultant hydrocarbon reaction product remains as one or more aliphatic hydrocarbons or as an alkyl group on an aromatic or naphthenic hydrocarbon.

Middle distillate is combined with recycle and makeup hydrogen, preheated to reaction temperature in a feed/effluent exchanger and a fired heater and charged to the first reactor (two or more reactors in series are required). In the reactors the contained sulphur, nitrogen and oxygen compounds are converted over the "Hydrobon" catalyst, with aromatics being saturated to improve middle distillate properties (as an acceptable FCC feedstock).

The reaction product is cooled via heat exchange and air cooling enroute to the high pressure product separator. Net separator gas is recycled and combined with hydrogen-rich makeup gas. The separator liquid is charged to a product fractionator, where H_2S and undesired light ends are stripped from the treated product. The bottoms product from the fractionator is cooled via heat exchange and air cooling and pumped to final product storage at a rate of 65.0 tonne/hr (11 020 barrels per day).

7.3.4.3 Area 700 - Heavy Distillate Hydrotreater

Heavy distillate, 350-525°C, from the fractionator bottoms and vacuum tower distillate at a rate of 38.7 tonne/hr is processed in the Heavy Distillate Hydrotreater to yield a stable heavy distillate product. Low in sulphur, nitrogen and oxygen, this liquid is suitable for use of a heavy boiler fuel or as Fluid Catalytic Cracker feedstock. Critical product specifications are <2 000 ppm nitrogen content, <100 ppm sulphur content and <2 000 ppm oxygen content.

The UOP HD Unibon Process, similar to that used in the Middle Distillate Hydrotreating Unit was selected since UOP Process Division experience includes treating of heavy distillates as well as middle distillates. Their recent pilot plant studies have included heavy coal-derived oils.

The severity of hydrotreating in the Heavy Distillate Hydrotreater is higher than in the Middle Distillate Hydrotreater in order to meet the required critical product specifications. As a result, moderate hydrocracking occurs yielding partially hydrotreated naphtha and middle distillate streams.

Heavy Distillate is combined with recycle and makeup hydrogen, preheated to reaction temperature and charged to the first reactor (two or more reactors in series are required). In the reactors sulphur, nitrogen and oxygen compounds are converted over the Hydrobon catalyst to hydrogen sulphide, ammonia and water respectively.

The reaction product is cooled via heat exchange and air cooling en route to the high pressure product separator. Net separator gas is recycled and combined with hydrogen rich make-up gas. The separator liquid is routed to a fractionator, where H_2S, NH_3 and light ends (C_1-C_4) are taken overhead and routed to the ammonia recovery unit. Naphtha and middle distillate side cuts are taken from the fractionator for final treatment in the naphtha and middle distillate hydrotreaters. A small fraction of the naphtha stream is used as make-up for the Phenol Recovery Unit. The bottoms product of the fractionator is cooled and sent to storage at a rate of 26.8 tonne/hr (4 215 barrels per day).

7.3.5 Area 800 - Light Ends Recovery

The sweet gases from the Coal Liquefaction Unit, Area 300, and the Stretford Sulphur Recovery Unit, Area 900, are processed in this area.

7.3.5.1 Carbon Dioxide Removal

Sweet hydrocarbon gases from the sulphur plant are fed to an amine unit for CO_2 removal. This unit uses diglycolamine to combine with CO_2 in the amine absorber. The spent amine is withdrawn from the bottom of the absorber and sent to an amine stripper where the CO_2 is released and vented to the atmosphere. The regenerated diglycolamine is cooled and recycled to the absorber. The CO_2 free gas is then processed in the LPG recovery unit.

7.3.5.2 LPG Recovery

Feedstock to this unit is sweet gas from the coal liquefaction cryogenic unit and the CO_2 removal unit. The LPG feed is contacted with lean oil (naphtha) in a trayed absorber. Hydrocarbons are absorbed in the lean oil and exit via the absorber bottoms. Overhead gases from the absorber are routed to a cryogenic unit while the absorber bottoms are fed to a debutanizer.

Propane and butane are stripped from the naphtha in the debutanizer and are fed to a depropanizer. Stripped naphtha or lean oil is recycled back to the absorber feed to close the lean oil loop. A portion of the lean oil stream, representing the amount of naphtha recovered from the feed streams, is withdrawn and sent to naphtha hydrotreater. Propane and butane are separated into separate components in the depropanizer and sent to finished product storage as 12.4 tonne/hr of propane and 11.3 tonne/hr of butane. The depropanizer and debutanizer towers are heat integrated for maximum energy conservation.

7.3.5.3 Cryogenic Unit

The overhead gas from the absorber tower in the LPG Recovery Plant is routed to the Cryogenic Unit to maximize the recovery of hydrogen for recycling to the Coal Liquefaction Area.

The feed stream is processed through molecular sieve adsorbers to remove moisture and CO_2 in order to prevent blockages in the cryogenic heat exchangers. The moisture and carbon dioxide-free gas passes through a series of heat exchangers and cascading Joule-Thomson expansions to achieve the necessary refrigeration. Heavier hydrocarbon components are condensed, revaporized and distributed as plant fuel gas. The cold, hydrogen-rich stream leaves the unit countercurrently through the same heat exchangers, cooling the incoming gas, and is sent to a PSA unit for high purity hydrogen recovery.

7.3.5.4 PSA Unit

The PSA unit processes 91.2 Vol.% H_2 effluent gas from the Cryogenic unit and produces 99.5 Vol.% hydrogen which is recompressed to 21 000 kPa and recycled to the Coal Liquefaction Area.

The unit uses a 4 bed adsorption system, with one bed on adsorption and 3 beds on various stages of regeneration. Purge gas is passed through the on-stream adsorption bed where the C_1-C_5 hydrocarbons, water, carbon monoxide and carbon dioxide molecules are adsorbed. The contained hydrogen passes through the adsorption bed and becomes part of the recovered hydrogen stream. When the adsorption capacity of the bed is depleted, the purge gas is switched automatically to a regenerated adsorber and the depleted adsorber is regenerated.

Regeneration of the adsorption bed is accomplished in three stages-depressurization, stripping with recovered hydrogen and repressurization to adsorption pressure. During the depressurization and stripping stages, the C_1-C_5 hydrocarbons, carbon monoxide and carbon dioxide are desorbed, discharged at 170 kPa, recompressed to 2 340 kPa and recycled to the Cryogenic Unit.

7.3.6 Area 900 - Sulphur Plant

Sour gas streams from the coal conversion units, product separation, product upgrading, residue gasifiers and ammonia recovery are treated in the Sulphur Plant to remove hydrogen sulphide and produce a sweet hydrocarbon stream as feed to the Light Ends Recovery. The Stretford Process was selected for this operation as it is a proven method of removing hydrogen sulphide from hydrocarbon gas streams and converting it directly to elemental sulphur. It offers both economic and operational advantages over the conventional Amine H_2S Scrubbing System plus Claus Sulphur Plant, particularly for the treatment of hydrocarbon gas streams with low H_2S content and high CO_2 content.

In the Stretford Unit, the sour fuel gas is countercurrently scrubbed in a venturi-absorber with an alkaline solution containing a vanadium salt along with anthraquinonedisulfonic acid (ADA). In a single step, the H_2S is converted into elemental sulphur by the vanadic salt which itself is reduced to the vanadous form, from the pentavalent to the quadrivalent state.

Sweet hydrocarbon gas is withdrawn from the absorber and fed to the CO_2 removal unit in the light ends recovery area.

The reduced liquor from the absorber is fed to the oxidizer where the vanadium is restored to its vanadic form via a redox mechanism with ADA. Air is blown through the liquor to reoxidize the ADA and separate the sulphur by froth flotation. The froth is either filtered or centrifuged, washed and melted to produce high-quality sulphur.

Due to the formation of by-product salts, such as sodium thiosulphate and sodium thiocyanate, it is necessary to purge liquor from the sulphur recovery section of the Stretford Unit. A proprietary reductive incineration process is included to reduce the purge stream to hydrogen sulphide and sodium carbonate, which are recycled to the front end of the plant, thereby resulting in a zero effluent discharge from the plant.

Liquid sulphur from the Sulphur Recovery Section is cooled, slated in a conventional Sulphur Slating Unit and stockpiled. The sulphur product, 35 tonne per day, is bright yellow product, 99.8+ wt% sulphur, with less than 1 ppm H_2S content.

7.3.7 Area 1000 - Hydrogen Production

Conceptual Flow Diagram 100-43-004 depicts the major equipment in a Texaco Partial Oxidation Unit feeding residue. Information on the status of development of the Texaco Gasifier is given in Section 4.7 of this report.

7.3.7.1 Gasification

The residue from the vacuum tower is pumped to a day storage tank. The contents of the day tank are kept fluid at about 300°C by constant recirculation through a fired heater. The residue solidifies at around 200°C.

The fluid residue is pumped to the surge tanks of each of the gasifier trains to be fed to the gasifiers. The residue in the surge tanks is also recirculated via the fired heater to prevent solidification.

The gasification units also have the flexibility to utilize the stored solid residue. The solid residue is granulated, as described in Section 7.3.3, ground and conveyed to a preheating vessel. Here the finely ground residue is heated and melted before transferring to the day tank.

The residue reacts in the refractory-lined upper chamber of the gasifier with steam and sub-stoichiometric oxygen to form a gaseous product consisting primarily of hydrogen and carbon monoxide with lesser amounts of water vapour, carbon dioxide, hydrogen sulphide and/or carbonyl sulphide, methane and nitrogen. Because of the high gasifier temperature, no tars, phenols or other undesirable by-products are produced. The per pass carbon conversion is primarily controlled by the amount of oxygen fed to the gasifier with the residue. Conversion in excess of 90% can be expected. Ungasified carbon is recycled to the gasifier in order to increase the overall conversion.

The gas from the oxidation chamber passes straight down into the quench tank at the bottom of the gasifier. Here the molten ash is immediately quenched with water and agglomerates into essentially carbon-free molten slag droplets. The slag droplets solidify and settle into the lock hopper. The lock hopper is periodically dumped onto a screen, from which the slag is conveyed to the solids treatment system.

The particulates in the gas stream are removed in cyclone separators and the gas is quenched in a quench chamber. The quenching of the gas further facilitates removal of any remaining solid particles. The clean gas is then fed to the venturi scrubber where the water and gas are separated. The scrubbed gas is saturated with water at approximately 200°C and is sent to the shift conversion unit to upgrade the hydrogen content.

The solids from the cyclone separators are conveyed back to the gasifier by the recycled raw gas via an eductor.

The wet synthesis gas from the venturi scrubbers are combined and undergo sour shifting. Because of the presence of sufficient water vapour in the synthesis gas, line steam injection is not required to offset the CO shift reaction:

$$CO + H_2O \longrightarrow CO_2 + H_2$$

The sour shift conversion reaction occurs over a conventional fixed bed cobalt/molybdenum catalyst. The reaction is highly exothermic and substantial steam is generated. It is assumed that the shift configuration employs feed/effluent exchangers to provide for preheating the feed gas to the desired temperature level. The overall reactor configuration may involve multiple stage reactors with interstage cooling/waste heat steam generation.

To maintain the plant operating 310 days per year, a total of three gasifiers are provided, two of which are normally operating with one standby. Commercial experience with other types of gasifiers has shown that, due to high maintenance, spare gasifiers are necessary to maintain consistent high operating factors.

7.3.7.2 Low Pressure Sour Gas Treatment

To recover pure hydrogen from the gasification and shift reactions, H_2S, CO_2 and COS must be removed. Several different designs of acid gas removal and sulphur recovery plants are available. The type selected depends on may factors including economics and coal sulphur content. For this study, a Selexol unit was chosen to treat the raw product gases from the gasifiers as the raw gases contain high CO_2 and low H_2S concentrations.

The Selexol unit removes H_2S, CO_2 and COS from the raw gas by means of a physical solvent, dimethyl ether of polyethylene glycol (Selexol). After absorption of the gaseous impurities, the Selexol solution is regenerated and recycled for further absorption. Selexol can be used selectively to enhance CO_2 absorption over H_2S. This permits concentration of H_2S to sufficient levels for recovery of the sulphur in the downstream sulphur plant.

The raw gas is cooled by heat exchange with the treated gas. Cooling the gas serves a dual purpose. Selexol is capable of absorbing more acid gas when operated at lower temperatures. Also, any water in the raw gas is condensed, and is removed before the cooled gas enters the absorption tower at the bottom. Cold lean Selexol enters the tower at the top and contacts the gas countercurrently.

The rich Selexol flows by gravity to a sump tank which provides a few minutes hold time to allow the entrained hydrogen to separate and be returned to the inlet stream. A small in-line centrifugal blower is used to return the hydrogen vapours.

From the sump tank, rich Selexol flows to a solvent heater where it is heated to the temperature level sufficient for H_2S and CO_2 recovery. The heated solution is then flashed to successively lower pressures in three stages, recovering as much energy as possible in the process. The flash gases from the intermediate stage, rich in CO_2, are vented.

The lean Selexol is cooled and recycled to the absorption tower. Flash gases from the high pressure and low pressure flash stages, rich in hydrogen, H_2S and CO_2 are then directed to the Stretford Sulphur Recovery Unit, where the H_2S is removed prior to hydrogen recovery and eventual feed to the coal liquefaction unit.

7.3.8 Area 1100 - Oxygen Plant

7.3.8.1 Oxygen Plant

The oxygen for the Texaco gasifiers is produced by a standard cryogenic low pressure air separation plant.

After filtration, compression and drying, the air is expanded to generate the cryogenic temperatures (approximately -170°C) required for liquefaction of nitrogen and oxygen. A rectification column separates these constituents to achieve the required oxygen purity of 98.5 wt%. The purified nitrogen and oxygen products are recovered as low pressure gases and must be compressed separately as required for in-plant use. The compressors have been specified with gas turbine drivers.

Two 50% capacity oxygen plants are provided to minimize plant upsets in the event of unscheduled shutdowns.

7.3.9 Area 1300 - Phenol Recovery

Sour water from coal conversion units, product separation, product upgrading and hydrogen production is treated in a Phenol Recovery unit (developed by Koppers Co. of Pittsburgh, Pennsylvania) and then transferred to an Ammonia Recovery unit. The unit removes 99% of the phenolics which are incinerated in one of the fired heaters. In the process, sour water is collected in a storage tank and pumped to the top of the dephenolizing tower. It then flows downward, countercurrent to an upward flow of solvent (stabilized naphtha). The multistage liquid-liquid extraction results in extraction of nearly all the phenolics from the sour water by the solvent. Dephenolized water drains from the bottom of the tower to the Ammonia Recovery unit and the phenolized solvent exits the top of the dephenolizing tower and flows to the solvent separator stage of the second tower. Entrained sour water is separated from the solvent, drained from the separator periodically and recycled.

Solvent overflows the separator and enters the first (lower) caustic washer and passes upward through a 25% caustic solution in a packed bed where the phenol reacts to form sodium phenolate. The solvent then exits the first caustic washer and enters the second caustic washer to remove the final traces of phenol. Dephenolized solvent overflows the second caustic washer and drains to the solvent circulating tank at the base of the second tower. Solvent is then recycled from the circulating tank back to the base of the dephenolizing tower.

Sodium phenolate is periodically drained from the lower caustic washer into the sodium phenolate storage tank. Caustic from the top caustic washer is drained into the lower caustic washer and fresh caustic is charged to the upper bed. Sodium phenolate is periodically transferred from the storage tank to the springing tank where excess water is evaporated before the phenolate is converted into phenol by CO_2 sparging.

After sparging, the solution is allowed to separate. Phenol is drained to the phenol storage tank and sodium carbonate solution to be sold as a by-product.

7.3.10 Area 1400 - Ammonia Recovery

Sour water from the Phenol Recovery unit is transferred to the Ammonia Recovery unit and is fed to a sour water WWT process unit developed by Chevron Research Co. of San Francisco, California. The unit removes hydrogen sulphide as a gas and anhydrous ammonia and the resulting stream contains only 50 mg/L ammonia and 5 mg/L hydrogen sulphide.

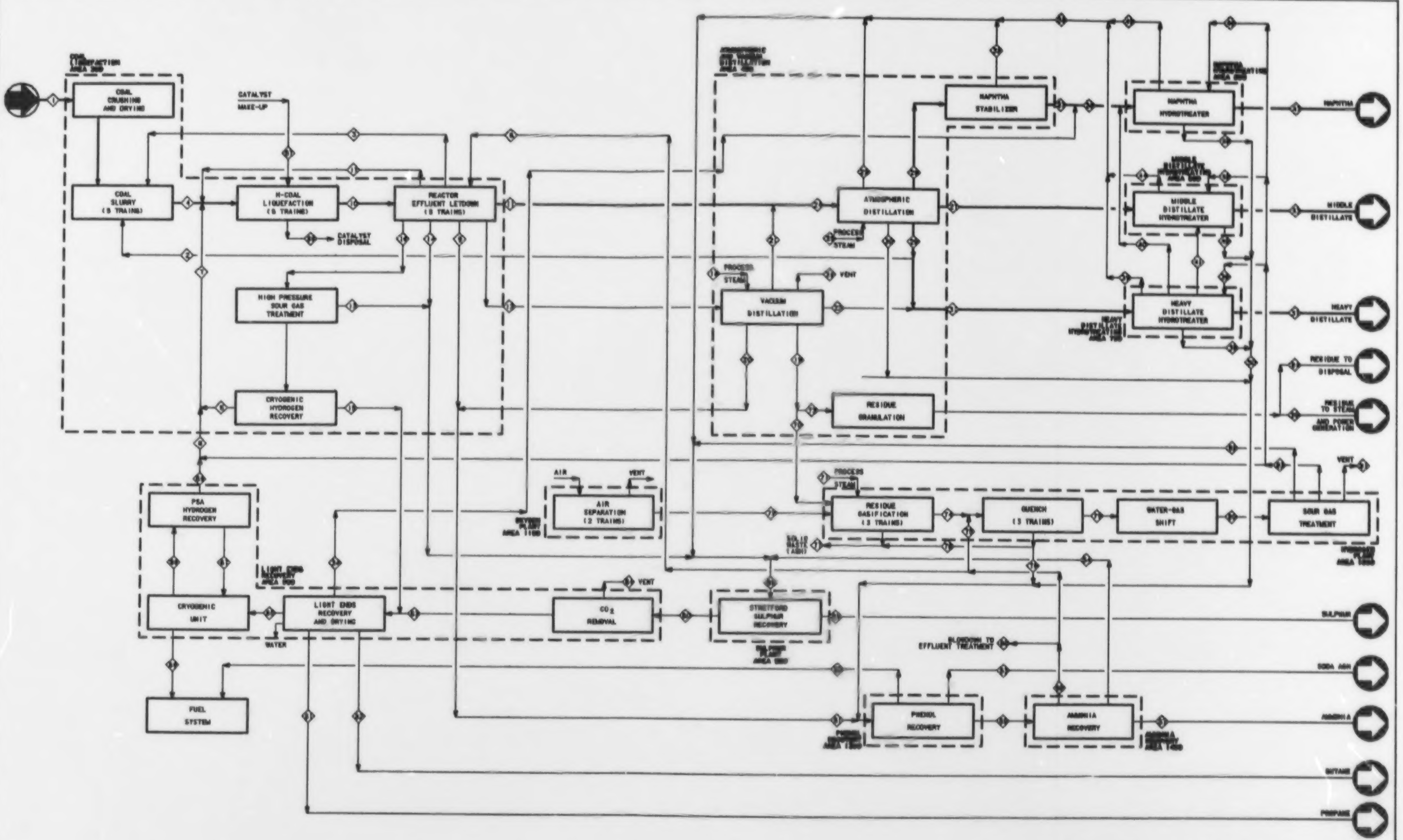
In the process, sour water from the Phenol Recovery unit is combined with recycle water from the Ammonia Purification unit, cooled and passed through a degasser where dissolved hydrocarbons are removed. The recycle stream is rich in ammonia which minimizes the release of acid gas at this point. The degassed water is then pumped to storage to allow for stabilization in flow rates and separation of entrained solids and oil.

Degassed sour water is pumped to the hydrogen sulphide stripper. Hydrogen sulphide is stripped overhead and water washed to reduce ammonia contamination. The resulting gas stream is blended with offgas from the degasser and sent to the Sulphur Plant.

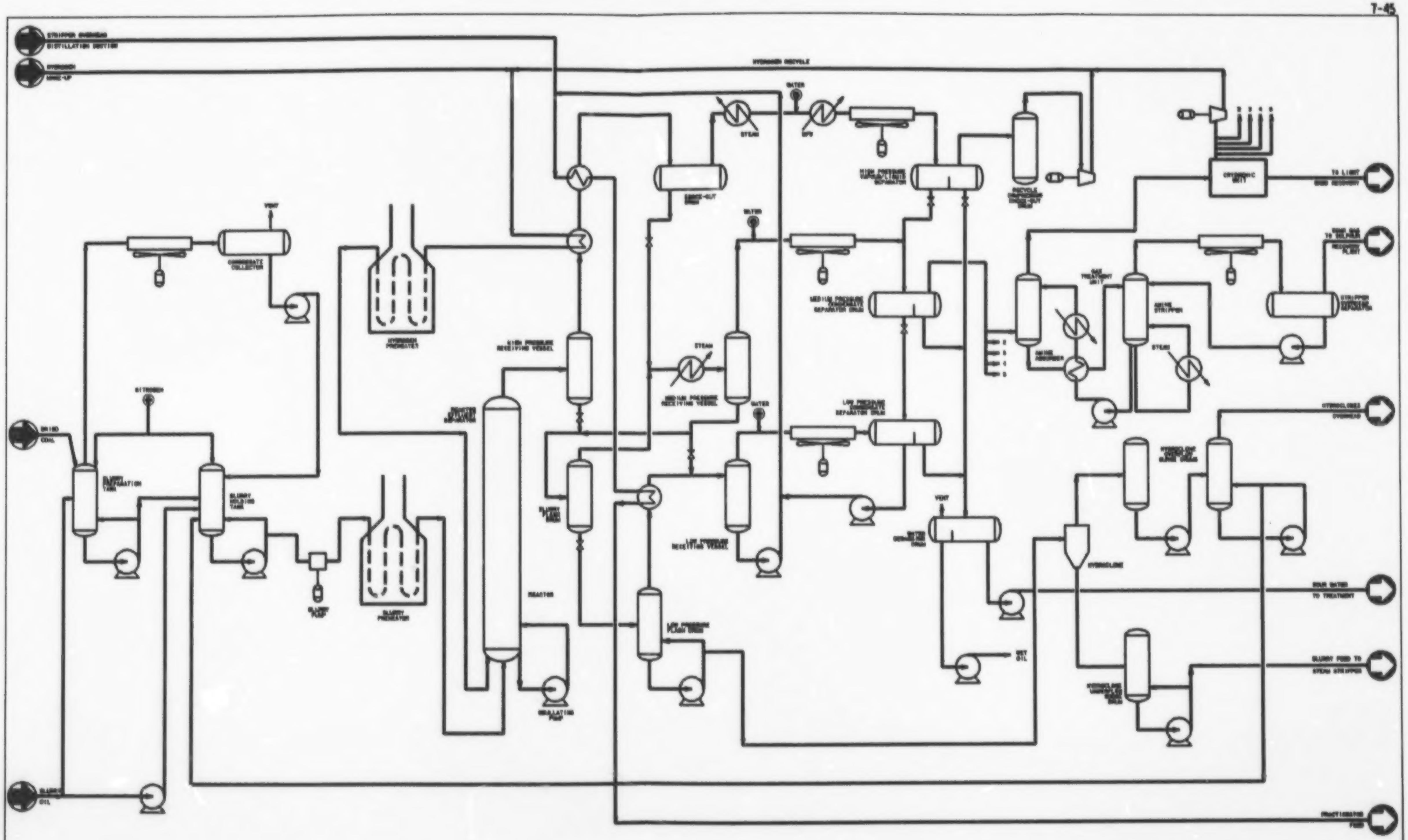
The hydrogen sulphide stripped bottoms containing ammonia and some hydrogen sulphide is fed directly to the ammonia stripper. The ammonia stripper is a steam reboiled refluxed column where essentially all the ammonia and remaining hydrogen sulphide is removed from the water which leaves as the column bottoms. After cross-exchanging with the hydrogen sulphide stripper feed and cooling, the water is transferred to effluent treatment.

Ammonia and hydrogen sulphide stripped from the water in the ammonia stripper is partially condensed. The liquid is recycled and the gas is scrubbed in a two stage scrubber to remove hydrogen sulphide. The scrubber liquid is recycled to the degasser and the vapour is compressed and liquefied to produce 4.8 tonne/hr of anhydrous ammonia. The ammonia is transferred to the product tankage area for sale as a by-product.

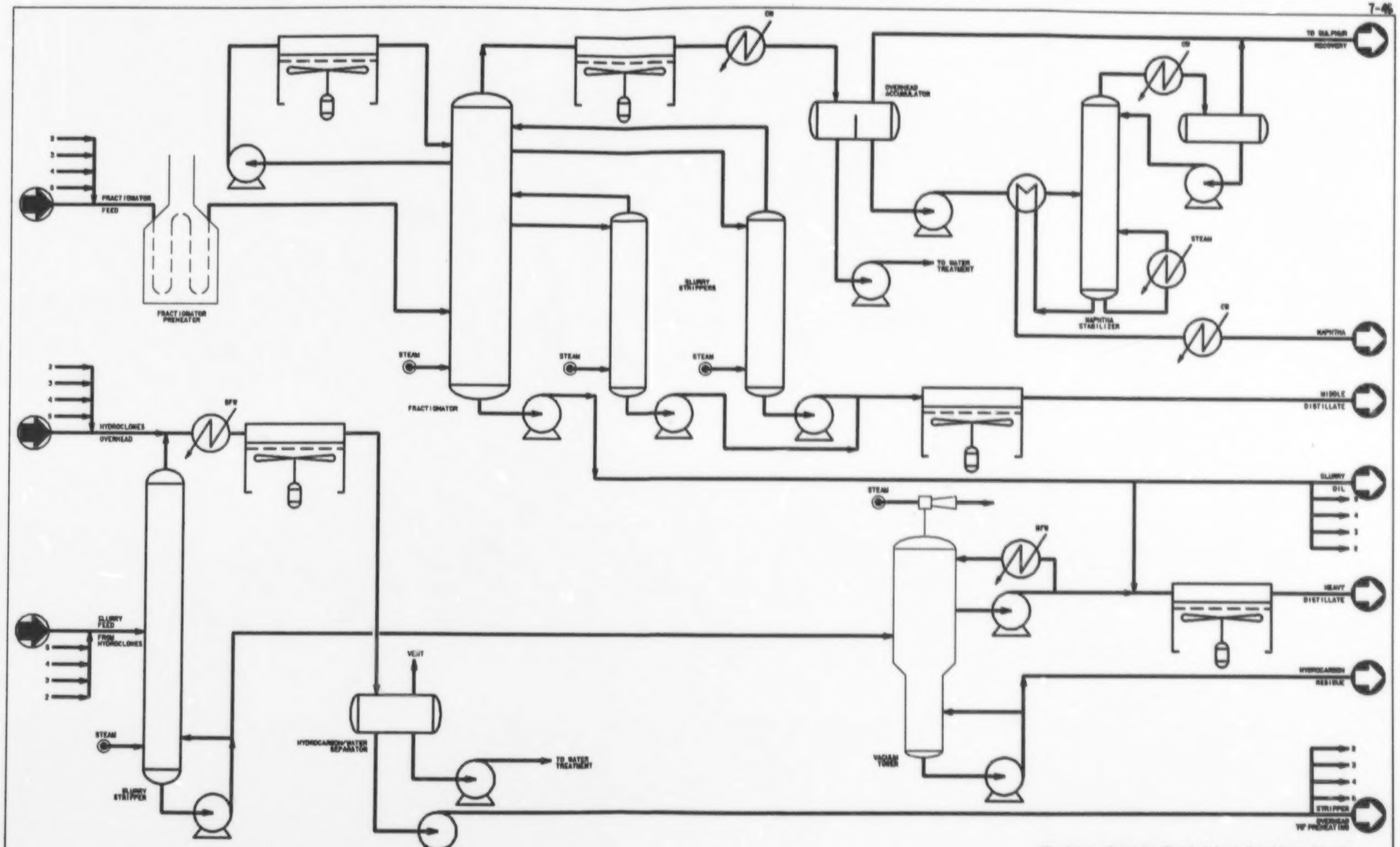
7.4 BLOCK FLOW DIAGRAMS


COAL LIQUEFACTION AREA 200
COAL CRUSHING AND DRYING
CATALYST MAKE-UP
COAL SLURRY (5 TRAINS)
H-COAL LIQUEFACTION (5 TRAINS)
CATALYST DISPOSAL
REACTOR EFFLUENT LETDOWN (5 TRAINS)
HIGH PRESSURE SOUR GAS TREATMENT
CRYOGENIC HYDROGEN RECOVERY
ATMOSPHERIC AND VACUUM DISTILLATION AREA 400
NAPHTHA STABILIZER
ATMOSPHERIC DISTILLATION
PROCESS STEAM
VACUUM DISTILLATION
VENT
RESIDUE GRANULATION
NAPHTHA HYDROTREATER
MIDDLE DISTILLATE HYDROTREATER
HEAVY DISTILLATE HYDROTREATER
HEAVY DISTILLATE HYDROTREATING AREA 700
NAPHTHA
MIDDLE DISTILLATE
HEAVY DISTILLATE
RESIDUE TO DISPOSAL
RESIDUE TO STEAM AND POWER GENERATION
PSA HYDROGEN RECOVERY
LIGHT ENDS RECOVERY AREA 900
CRYOGENIC UNIT
LIGHT ENDS RECOVERY AND DRYING
WATER
FUEL SYSTEM
AIR
VENT
AIR SEPARATION (2 TRAINS)
OXYGEN PLANT AREA 1100
CO₂ REMOVAL
VENT
STRETFORD SULPHUR RECOVERY
SULPHUR PLANT AREA 1300
PROCESS STEAM
RESIDUE GASIFICATION (3 TRAINS)
SOLID WASTE (ASH)
QUENCH (3 TRAINS)
WATER-GAS SHIFT
SOUR GAS TREATMENT
VENT
HYDROGEN PLANT AREA 1000
BLOWDOWN TO EFFLUENT TREATMENT
PHENOL RECOVERY
PHENOL RECOVERY AREA 1300
AMMONIA RECOVERY
AMMONIA RECOVERY AREA 1400
SULPHUR
SODA ASH
AMMONIA
BUTANE
PROPANE

Western Canada Coal Liquefaction Study
Prefeasibility Study
Simplified Block Flow Diagram
H-Coal Base Case
Drawing No. 100-43-001


STRIPPER OVERHEAD
DISTILLATION SECTION
HYDROGEN
MAKE-UP
HYDROGEN RECYCLE
WATER
STEAM
BFW
VENT
CONDENSATE COLLECTOR
HYDROGEN PREHEATER
HIGH PRESSURE RECEIVING VESSEL
KNOCK-OUT DRUM
HIGH PRESSURE VAPOR/LIQUID SEPARATOR
RECYCLE COMPRESSOR KNOCK-OUT DRUM
CRYOGENIC UNIT
TO LIGHT
ENDS RECOVERY
SOUR GAS TO SULPHUR
RECOVERY PLANT
MEDIUM PRESSURE CONDENSATE SEPARATOR DRUM
GAS TREATMENT UNIT
AMINE ABSORBER
AMINE STRIPPER
STRIPPER OVERHEAD SEPARATOR
MEDIUM PRESSURE RECEIVING VESSEL
NITROGEN
DRIED
COAL
SLURRY PREPARATION TANK
SLURRY HOLDING TANK
SLURRY PUMP
SLURRY PREHEATER
REACTOR
CIRCULATING PUMP
SLURRY FLASH DRUM
LOW PRESSURE CONDENSATE SEPARATOR DRUM
LOW PRESSURE RECEIVING VESSEL
LOW PRESSURE FLASH DRUM
HYDROCLONES
OVERHEAD
HYDROCLONE
WET
OIL
SOUR WATER
TO TREATMENT
SLURRY FEED TO
STEAM STRIPPER
FRACTIONATOR
FEED
SLURRY
OIL

Western Canada Coal Liquefaction Study
Prefeasibility Study
Conceptual FLow Sheet
Coal Conversion Unit
Drawing No. 100-43-002


2
3
4
5
FRACTIONATOR
FEED
FRACTIONATOR
PREHEATER
CW
OVERHEAD
ACCUMULATOR
TO SULPHUR
RECOVERY
CW
TO WATER
TREATMENT
SLURRY
STRIPPERS
STEAM
NAPHTHA
STABILIZER
CW
NAPHTHA
STEAM
STEAM
STEAM
FRACTIONATOR
MIDDLE
DISTILLATE
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HYDROCLONES
OVERHEAD
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SLURRY
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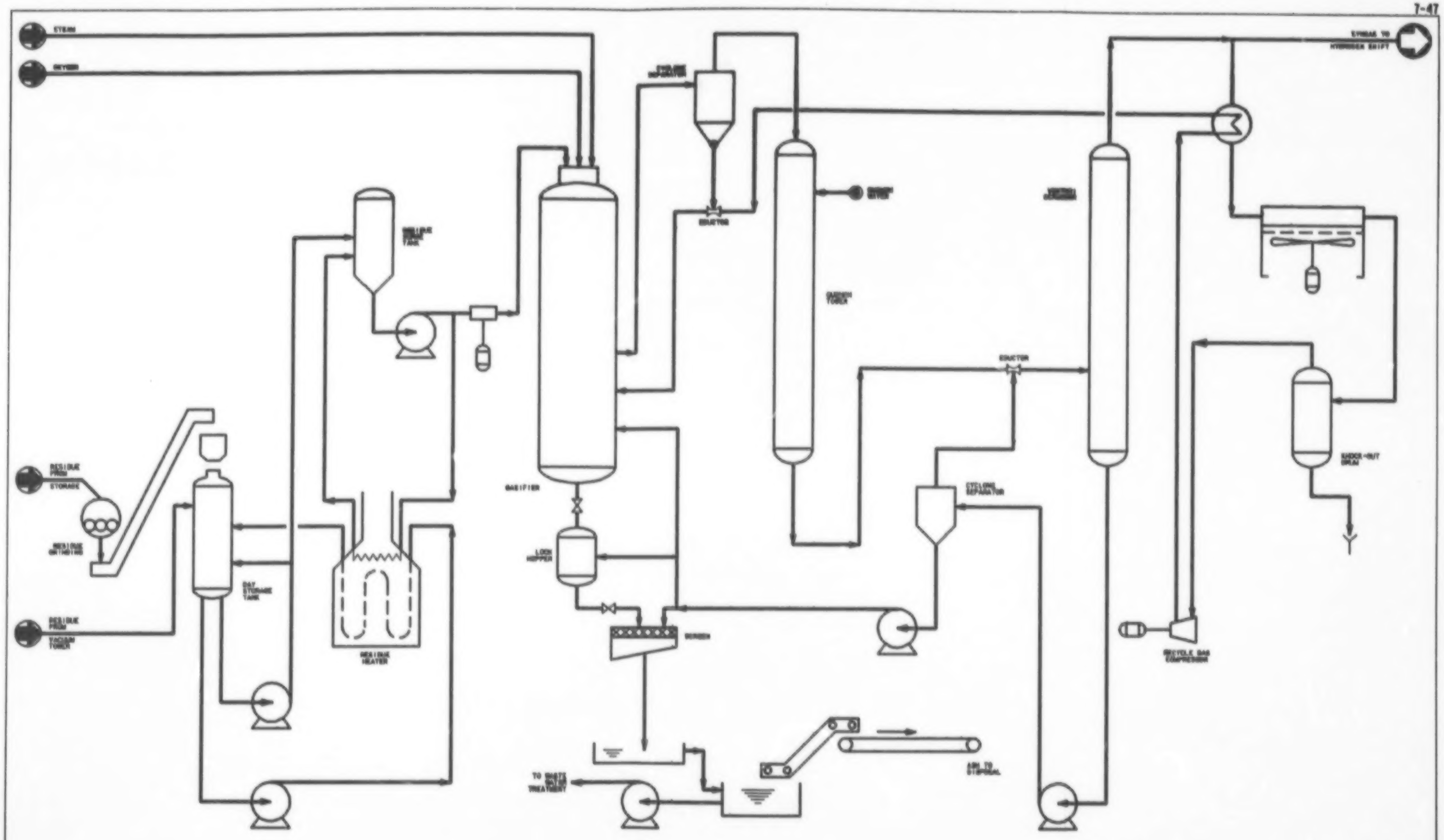
STEAM
BFW
HEAVY
DISTILLATE
SLURRY
FEED
FROM
HYDROCLONES
5
4
3
2
VENT
HYDROCARBON
RESIDUE
STEAM
HYDROCARBON/WATER
SEPARATOR
TO WATER
TREATMENT
VACUUM
TOWER
SLURRY
STRIPPER
2
3
4
5
STRIPPER
OVERHEAD
TO PREHEATING

Western Canada Coal Liquefaction Study
Prefeasibility Study
Conceptual Flow Sheet
Product Separation
Drawing No. [illegible]00-43-003


STEAM
OXYGEN
CYCLONE SEPARATOR
EDUCTOR
SYNGAS TO HYDROGEN SHIFT
RESIDUE FROM STORAGE
RESIDUE GRINDING
RESIDUE FROM VACUUM TOWER
DAY STORAGE TANK
RESIDUE HEATER
GASIFIER
LOCK HOPPER
SCREEN
EDUCTOR
CYCLONE SEPARATOR
KNOCK-OUT DRUM
RECYCLE GAS COMPRESSOR
TO WASTE WATER TREATMENT
ASH TO DISPOSAL

Western Canada Coal Liquefaction Study
Prefeasibility Study
Conceptual Flow Sheet
Residue Gasification
Drawing No. 100-43-004

7.5 MATERIAL BALANCE H-COAL BASE CASE

"H-COAL" - MATERIAL BALANCE

STREAM NO.	1	2	3
STREAM NAME	Coal	Slurry Oil	Hydroclone Recycle
TOTAL FLOW RATE:			
Kg/hr	416666	694489	136386
BBL/SD			
TEMPERATURE °C	40	290	400
PRESSURE kPa	100	720	2600
COMPONENT FLOW RATE:			
C_1			3
C_2			5
C_3			5
C_4			7
C_5 to 200°C		1080	192
200°C to 350°C		34611	767
350°C to 525°C		657660	129405
Solid hydrocarbon residue			4910
Coal feedstock (maf)	350000		
Unreacted coal (maf)			545
Ash	33333		491
H_2			2
H_2O	33333	1138	40
CO			2
CO_2			5
NH_3			2
H_2S			5
SO_2			
N_2			
Oxygen			

"H-COAL" - MATERIAL BALANCE

STREAM NO.	4	5	6
STREAM NAME	Coal Slurry	Cryo H_2 Recycle	Hydrogen Makeup
TOTAL FLOW RATE:			
Kg/hr	1222031	29027	31512
BBL/SD			
TEMPERATURE °C	250	105	105
PRESSURE kPa	21750	21750	21750
COMPONENT FLOW RATE:			
C_1	3	2200	4430
C_2	5		
C_3	5		
C_4	7		
C_5 to 200°C	1272		
200°C to 350°C	35378		
350°C to 525°C	787065		
Solid hydrocarbon residue	4910		
Coal feedstock (maf)	350000		
Unreacted coal (maf)	545		
Ash	33824		
H_2	2	18114	20245
H_2O	9001	175	
CO	2	8538	2773
CO_2	5		4064
NH_3	2		
H_2S	5		
SO_2			
N_2			
Oxygen			

"H-COAL" - MATERIAL BALANCE

STREAM NO.	7	8	9
STREAM NAME	Hydrogen Feed	Quench Water	Sour Water
TOTAL FLOW RATE:			
Kg/hr	60539	224420	253336
BBL/SD			
TEMPERATURE °C	105	65	55
PRESSURE kPa	21750	19700	200
COMPONENT FLOW RATE:			
C_1	6630		
C_2			
C_3			
C_4			
C_5 to 200°C			
200°C to 350°C			
350°C to 525°C			
Solid hydrocarbon residue			
Coal feedstock (maf)			
Unreacted coal (maf)			
Ash			
H_2	38359		
H_2O	175	224420	249171
CO	11311		
CO_2	4064		430
NH_3			3505
H_2S			230
SO_2			
N_2			
Oxygen			

"H-COAL" - MATERIAL BALANCE

STREAM NO.	10	11	12
STREAM NAME	Reactor Effluent	Distillation Feed	Residue Slurry
TOTAL FLOW RATE:			
Kg/hr	1354362	632020	380767
BBL/SD			
TEMPERATURE °C	Confidential	310	405
PRESSURE kPa	Confidential	620	200
COMPONENT FLOW RATE:			
C_1	51531	54	3
C_2	20606	193	10
C_3	16338	446	10
C_4	9735	1095	22
C_5 to 200°C	43922	30984	2100
200°C to 350°C	97118	86644	9707
350°C to 525°C	834466	511554	193507
Solid hydrocarbon residue	124671		119761
Coal feedstock (maf)			
Unreacted coal (maf)	22701		22156
Ash	33824		33333
H_2	37731		
H_2O	28035	683	100
CO	20466	10	2
CO_2	8392	82	15
NH_3	3572	45	20
H_2S	1254	230	21
SO_2			
N_2			
Oxygen			

"H-COAL" - MATERIAL BALANCE

STREAM NO.	13	14	15
STREAM NAME	Sour Gas (From H.P. Amine)	Sour Gas (From L.P.Letdown)	Sweet Gas (Co.Liq. H_2 Cryo)
TOTAL FLOW RATE:			
Kg/hr	5969	18552	50933
BBL/SD			
TEMPERATURE °C	45	50	5
PRESSURE kPa	160	330	330
COMPONENT FLOW RATE:			
C_1	6	1878	16673
C_2	8	2203	11155
C_3	6	2859	9669
C_4	4	3298	4918
C_5 to 200°C		5109	4887
200°C to 350°C			
350°C to 525°C			
Solid hydrocarbon residue			
Coal feedstock (maf)			
Unreacted coal (maf)			
Ash			
H_2			1501
H_2O	1116	850	320
CO		1566	1810
CO_2	4315	635	
NH_3		-	
H_2S	514	154	
SO_2			
N_2			
Oxygen			

"H-COAL" - MATERIAL BALANCE

STREAM NO.	16	17	18
STREAM NAME	H.P. Amine Unit Feed	Recycle Hydrogen	Steam to Steam Stripper
TOTAL FLOW RATE:			
Kg/hr	85929	71792	5470
BBL/SD			
TEMPERATURE °C	50	50	180
PRESSURE kPa	8070	20200	1130
COMPONENT FLOW RATE:			
C_1	18879	30714	
C_2	11163	7032	
C_3	9675	3343	
C_4	4922	391	
C_5 to 200°C	4887	650	
200°C to 350°C			
350°C to 525°C			
Solid hydrocarbon residue			
Coal feedstock (maf)			
Unreacted coal (maf)			
Ash			
H_2	19615	18114	
H_2O	1611		5470
CO	10348	8538	
CO_2	4315	2910	
NH_3			
H_2S	514	100	
SO_2			
N_2			
Oxygen			

"H-COAL" - MATERIAL BALANCE

STREAM NO.	19	20	21
STREAM NAME	Vacuum Tower Bottoms	Sour Water	Steam Stripper OVHD Distillate
TOTAL FLOW RATE:			
Kg/hr	186371	5340	170869
BBL/SD			
TEMPERATURE °C	370	95	95
PRESSURE kPa	310	260	260
COMPONENT FLOW RATE:			
C_1			1
C_2			8
C_3			8
C_4			18
C_5 to 200°C			2040
200°C to 350°C			7303
350°C to 525°C	11121		161372
Solid hydrocarbon residue	119761		
Coal feedstock (maf)			
Unreacted coal (maf)	22156		
Ash	33333		
H_2			
H_2O		5320	100
CO			
CO_2			
NH_3		20	
H_2S			19
SO_2			
N_2			
Oxygen			

"H-COAL" - MATERIAL BALANCE

STREAM NO.	22	23	24
STREAM NAME	Vent	Vacuum Tower Distillate	Fractionator Feed
TOTAL FLOW RATE:			
Kg/hr	189	23468	822719
BBL/SD			
TEMPERATURE °C	95	210	310
PRESSURE kPa	100	400	620
COMPONENT FLOW RATE:			
C_1	2	-	55
C_2	2	-	201
C_3	2	-	454
C_4	4	-	1113
C_5 to 200°C	10	50	33024
200°C to 350°C		2404	93947
350°C to 525°C		21014	672926
Solid hydrocarbon residue			
Coal feedstock (maf)			
Unreacted coal (maf)			
Ash			
H_2			
H_2O	150		20613
CO	2		10
CO_2	15		82
NH_3			45
H_2S	2		249
SO_2			
N_2			
Oxygen			

"H-COAL" - MATERIAL BALANCE

STREAM NO.	25	26	27
STREAM NAME	Steam	Fractionator Bottoms	Fractionator Sidecut
TOTAL FLOW RATE:			
Kg/hr	19830	708627	59990
BBL/SD			
TEMPERATURE °C	180		
PRESSURE kPa	1130		
COMPONENT FLOW RATE:			
C_1			
C_2			
C_3			
C_4			
C_5 to 200°C		1105	669
200°C to 350°C		35414	58503
350°C to 525°C		672108	818
Solid hydrocarbon residue			
Coal feedstock (maf)			
Unreacted coal (maf)			
Ash			
H_2			
H_2O	19830		
CO			
CO_2			
NH_3			
H_2S			
SO_2			
N_2			
Oxygen			

"H-COAL" - MATERIAL BALANCE

STREAM NO.	28	29	30
STREAM NAME	Naphtha Stab. Feed	Offgas	Sour Water
TOTAL FLOW RATE:			
Kg/hr	32890	565	20647
BBL/SD			
TEMPERATURE °C			
PRESSURE kPa			
COMPONENT FLOW RATE:			
C_1	15	40	
C_2	71	130	
C_3	419	35	
C_4	1071	42	
C_5 to 200°C	31205	45	
200°C to 350°C	30		
350°C to 525°C			
Solid hydrocarbon residue			
Coal feedstock (maf)			
Unreacted coal (maf)			
Ash			
H_2			
H_2O	5	6	20602
CO	2	8	
CO_2	40	42	
NH_3			45
H_2S	32	217	
SO_2			
N_2			
Oxygen			

"H-COAL" - MATERIAL BALANCE

STREAM NO.	31	32	33
STREAM NAME	Hvy. Distillate Feed	Stabilizer Offgas	Stabilized Naphtha
TOTAL FLOW RATE:			
Kg/hr	38744	1338	31552
BBL/SD			
TEMPERATURE ^oC		40	
PRESSURE kPa		380	
COMPONENT FLOW RATE:			
C_1		15	
C_2		71	
C_3		419	
C_4		642	429
C_5 to 200^oC	75	112	31093
200^oC to 350^oC	3207		30
350^oC to 525^oC	35462		
Solid hydrocarbon residue			
Coal feedstock (maf)			
Unreacted coal (maf)			
Ash			
H_2			
H_2O		5	
CO		2	
CO_2		40	
NH_3			
H_2S		32	
SO_2			
N_2			
Oxygen			

"H-COAL" - MATERIAL BALANCE

STREAM NO.	34	35	36
STREAM NAME	Stabilized Naphtha	Naphtha Hydro. Feed	H_2 to Heavy Distillate
TOTAL FLOW RATE:			
Kg/hr	10173	41725	4974
BBL/SD			
TEMPERATURE °C			50
PRESSURE kPa			2200
COMPONENT FLOW RATE:			
C_1			699
C_2			
C_3			
C_4	210	639	
C_5 to 200°C	9963	41056	
200°C to 350°C		30	
350°C to 525°C			
Solid hydrocarbon residue			
Coal feedstock (maf)			
Unreacted coal (maf)			
Ash			
H_2			3196
H_2O			
CO			438
CO_2			641
NH_3			
H_2S			
SO_2			
N_2			
Oxygen			

"H-COAL" - MATERIAL BALANCE

STREAM NO.	37	38	39
STREAM NAME	Heavy Distillate	Sour Water	Sour Gas
TOTAL FLOW RATE:			
Kg/hr	26805	753	5383
BBL/SD			
TEMPERATURE °C	150	40	100
PRESSURE kPa	400	200	330
COMPONENT FLOW RATE:			
C_1			1325
C_2			743
C_3			640
C_4			1278
C_5 to 200°C			
200°C to 350°C	1285		
350°C to 525°C	25520		
Solid hydrocarbon residue			
Coal feedstock (maf)			
Unreacted coal (maf)			
Ash			
H_2			290
H_2O		464	
CO			438
CO_2			641
NH_3		289	
H_2S			28
SO_2			
N_2			
Oxygen			

"H-COAL" - MATERIAL BALANCE

STREAM NO.	40	41	42
STREAM NAME	Stabilized Naphtha	Middle Distillate	H_2 to Middle Distillate
TOTAL FLOW RATE:			
Kg/hr	4471	6306	2159
BBL/SD			
TEMPERATURE °C	40	40	50
PRESSURE kPa	650	760	2200
COMPONENT FLOW RATE:			
C_1			304
C_2			
C_3			
C_4	2		
C_5 to 200°C	4449	200	
200°C to 350°C	20	6056	
350°C to 525°C		50	
Solid hydrocarbon residue			
Coal feedstock (maf)			
Unreacted coal (maf)			
Ash			
H_2			1388
H_2O			
CO			189
CO_2			278
NH_3			
H_2S			
SO_2			
N_2			
Oxygen			

"H-COAL" - MATERIAL BALANCE

STREAM NO.	43	44	45
STREAM NAME	Middle Distillate	Sour Gas	Sour Water
TOTAL FLOW RATE:			
Kg/hr	64977	2397	1081
BBL/SD			
TEMPERATURE ^{o}C	50	100	40
PRESSURE kPa	400	330	200
COMPONENT FLOW RATE:			
C_1		503	
C_2		308	
C_3		312	
C_4		628	
C_5 to 200^{o}C	1326		
200^{o}C to 350^{o}C	63320		
350^{o}C to 525^{o}C	331		
Solid hydrocarbon residue			
Coal feedstock (maf)			
Unreacted coal (maf)			
Ash			
H_2		126	
H_2O			631
CO		189	
CO_2		278	
NH_3			450
H_2S		53	
SO_2			
N_2			
Oxygen			

"H-COAL" - MATERIAL BALANCE

STREAM NO.	46	47	48
STREAM NAME	H_2 to Naphtha Hydrotreater	Naphtha	Sour Water
TOTAL FLOW RATE:			
Kg/hr	697	45542	417
BBL/SD			
TEMPERATURE °C	50	40	40
PRESSURE kPa	2200	400	200
COMPONENT FLOW RATE:			
C_1	98		
C_2			
C_3			
C_4		725	
C_5 to 200°C		44767	
200°C to 350°C		50	
350°C to 525°C			
Solid hydrocarbon residue			
Coal feedstock (maf)			
Unreacted coal (maf)			
Ash			
H_2	448		
H_2O			285
CO	62		
CO_2	89		
NH_3			132
H_2S			
SO_2			
N_2			
Oxygen			

"H-COAL" - MATERIAL BALANCE

STREAM NO.	49	50	51
STREAM NAME	Sour Gas	Hydrotreater Sour Water	Sour Water
TOTAL FLOW RATE:			
Kg/hr	934	2251	258676
BBL/SD			
TEMPERATURE °C	100	40	55
PRESSURE kPa	330	200	200
COMPONENT FLOW RATE:			
C_1	113		
C_2	15		
C_3	10		
C_4	570		
C_5 to 200°C			
200°C to 350°C			
350°C to 525°C			
Solid hydrocarbon residue			
Coal feedstock (maf)			
Unreacted coal (maf)			
Ash			
H_2	41		
H_2O		1380	254491
CO	62		
CO_2	89		430
NH_3		871	3525
H_2S	34		230
SO_2			
N_2			
Oxygen			

"H-COAL" - MATERIAL BALANCE

STREAM NO.	52	53	54
STREAM NAME	Phenols Recovered	Soda Ash	Sour Gas
TOTAL FLOW RATE:			
Kg/hr	400	378	932
BBL/SD			
TEMPERATURE °C	40	40	50
PRESSURE kPa	450	450	330
COMPONENT FLOW RATE:			
C_1			
C_2			
C_3			
C_4			
C_5 to 200°C			
200°C to 350°C			
350°C to 525°C			
Solid hydrocarbon residue			
Coal feedstock (maf)			
Unreacted coal (maf)			
Ash			
H_2			
H_2O	10		275
CO			
CO_2			427
NH_3			
H_2S			230
Soda Ash		378	
N_2			
Phenols	390		

"H-COAL" - MATERIAL BALANCE

STREAM NO.	55	56	57
STREAM NAME	Ammonia Recovery	Effluent Water	Ammonia
TOTAL FLOW RATE:			
Kg/hr	298752	292990	4830
BBL/SD			
TEMPERATURE °C	50	40	30
PRESSURE kPa	330	330	800
COMPONENT FLOW RATE:			
C_1			
C_2			
C_3			
C_4			
C_5 to 200°C			
200°C to 350°C			
350°C to 525°C			
Solid hydrocarbon residue			
Coal feedstock (maf)			
Unreacted coal (maf)			
Ash			
H_2			
H_2O	293249	292974	
CO			
CO_2	430	3	
NH_3	4843	13	4830
H_2S	230		
SO_2			
N_2			
Oxygen			

"H-COAL" - MATERIAL BALANCE

STREAM NO.	58	59	60
STREAM NAME	Hydrotreater Sour Gas	Elemental Sulphur	Sweet Gas
TOTAL FLOW RATE:			
Kg/hr	8713	1439	37211
BBL/SD			
TEMPERATURE °C	100		50
PRESSURE kPa	330		330
COMPONENT FLOW RATE:			
C_1	1941		4083
C_2	1066		3478
C_3	961		4280
C_4	2476		6462
C_5 to 200°C			5266
200°C to 350°C			
350°C to 525°C			
Solid hydrocarbon residue			
Coal feedstock (maf)			
Unreacted coal (maf)			
Ash			
H_2	457		457
H_2O			2252
CO	689		3498
CO_2	1008		7435
NH_3			
H_2S	115	1439	
SO_2			
N_2			
Oxygen			

"H-COAL" - MATERIAL BALANCE

STREAM NO.	61	62	63
STREAM NAME	Propane	Butane	CO_2 Free Gas
TOTAL FLOW RATE:			
Kg/hr	12415	11285	28857
BBL/SD			
TEMPERATURE °C	40	40	45
PRESSURE kPa	2000	760	2400
COMPONENT FLOW RATE:			
C_1			4083
C_2	138		3478
C_3	12132	130	4280
C_4	145	10965	6462
C_5 to 200°C		190	5266
200°C to 350°C			
350°C to 525°C			
Solid hydrocarbon residue			
Coal feedstock (maf)			
Unreacted coal (maf)			
Ash			
H_2			457
H_2O			1320
CO			3498
CO_2			13
NH_3			
H_2S			
SO_2			
N_2			
Oxygen			

"H-COAL" - MATERIAL BALANCE

STREAM NO.	64	65	66
STREAM NAME	Vent	Cryo Unit Feed	PSA Feed
TOTAL FLOW RATE:			
Kg/hr	8354	44277	4712
BBL/SD			
TEMPERATURE °C	60	40	40
PRESSURE kPa	100	2400	2400
COMPONENT FLOW RATE:			
C_1		20756	992
C_2		14495	558
C_3		1687	285
C_4		60	29
C_5 to 200°C			
200°C to 350°C			
350°C to 525°C			
Solid hydrocarbon residue			
Coal feedstock (maf)			
Unreacted coal (maf)			
Ash			
H_2		1958	2260
H_2O	932		
CO		5308	588
CO_2	7422	13	
NH_3			
H_2S			
SO_2			
N_2			
Oxygen			

"H-COAL" - MATERIAL BALANCE

STREAM NO.	67	68	69
STREAM NAME	PSA Recycle	Fuel Gas	PSA Product
TOTAL FLOW RATE:			
Kg/hr	2715	42280	1997
BBL/SD			
TEMPERATURE °C	40	40	40
PRESSURE kPa	2400	375	2200
COMPONENT FLOW RATE:			
C_1	992	20756	
C_2	558	14495	
C_3	285	1687	
C_4	29	60	
C_5 to 200°C			
200°C to 350°C			
350°C to 525°C			
Solid hydrocarbon residue			
Coal feedstock (maf)			
Unreacted coal (maf)			
Ash			
H_2	382	80	1878
H_2O			
CO	469	5189	119
CO_2		13	
NH_3			
H_2S			
SO_2			
N_2			
Oxygen			

"H-COAL" - MATERIAL BALANCE

STREAM NO.	70	71	72
STREAM NAME	Residue to H_2 Gasifier	Steam to Gasifier	O_2
TOTAL FLOW RATE:			
Kg/hr	132323	55575	81291
BBL/SD			
TEMPERATURE °C			
PRESSURE kPa			
COMPONENT FLOW RATE:			
C_1			
C_2			
C_3			
C_4			
C_5 to 200°C			
200°C to 350°C			
350°C to 525°C	7896		
Solid hydrocarbon residue	85030		
Coal feedstock (maf)			
Unreacted coal (maf)	15731		
Ash	23666		
H_2			
H_2O		55575	
CO			
CO_2			
NH_3			
H_2S			
Soot			
N_2 + Ar			1594
Oxygen			79697

"H-COAL" - MATERIAL BALANCE

STREAM NO.	73	74	75
STREAM NAME	Residue to Granulation	Gasifier Effluent Gas	Quench Water
TOTAL FLOW RATE:			
Kg/hr	54048	245523	145284
BBL/SD			
TEMPERATURE °C			
PRESSURE kPa			
COMPONENT FLOW RATE:			
C_1		5734	
C_2			
C_3			
C_4			
C_5 to 200°C			
200°C to 350°C			
350°C to 525°C	3225		
Solid hydrocarbon residue	34731		
Coal feedstock (maf)			
Unreacted coal (maf)	6425		
Ash	9667		
H_2		9859	
H_2O		2646	145284
CO		207508	
CO_2		16132	
NH_3		402	
H_2S		267	
Soot		1381	
N_2 + Ar		1594	
Oxygen			

"H-COAL" - MATERIAL BALANCE

STREAM NO.	76	77	78
STREAM NAME	Soot	Solid Waste	Sour Water
TOTAL FLOW RATE:			
Kg/hr	1381	23666	17956
BBL/SD			
TEMPERATURE ^{o}C			
PRESSURE kPa			
COMPONENT FLOW RATE:			
C_1			
C_2			
C_3			
C_4			
C_5 to $200^{o}C$			
$200^{o}C$ to $350^{o}C$			
$350^{o}C$ to $525^{o}C$			
Solid hydrocarbon residue			
Coal feedstock (maf)			
Unreacted coal (maf)			
Ash		23666	
H_2			17554
H_2O			
CO			
CO_2			
NH_3			402
H_2S			
Soot	1381		
N_2 + Ar			
Oxygen			

"H-COAL" - MATERIAL BALANCE

STREAM NO.	79	80	81
STREAM NAME	Shift Feed	Syngas	Selexol Vent
TOTAL FLOW RATE:			
Kg/hr	372102	372102	332086
BBL/SD			
TEMPERATURE °C			
PRESSURE kPa			
COMPONENT FLOW RATE:			
C_1	5734	5734	
C_2			
C_3			
C_4			
C_5 to 200°C			
200°C to 350°C			
350°C to 525°C			
Solid hydrocarbon residue			
Coal feedstock (maf)			
Unreacted coal (maf)			
Ash			
H_2	9859	23399	
H_2O	130376	-	
CO	207508	4700	124
CO_2	16764	336408	330368
NH_3			
H_2S	267	267	
SO_2			
N_2 + Ar	1594	1594	1594
Oxygen			

"H-COAL" - MATERIAL BALANCE

STREAM NO.	82	83	84
STREAM NAME	Acid Gas	H_2 Feed to Complex	Excess Residue
TOTAL FLOW RATE:			
Kg/hr	2671	37345	31143
BBL/SD			
TEMPERATURE °C			20
PRESSURE kPa			
COMPONENT FLOW RATE:			
C_1	203	5531	
C_2			
C_3			
C_4			
C_5 to 200°C			
200°C to 350°C			
350°C to 525°C			1859
Solid hydrocarbon residue			20011
Coal feedstock (maf)			
Unreacted coal (maf)			3703
Ash			5570
H_2		23399	
H_2O			
CO	1233	3343	
CO_2	968	5072	
NH_3			
H_2S	267		
SO_2			
N_2 + Ar			
Oxygen			

"H-COAL" - MATERIAL BALANCE

STREAM NO.	85	86	87
STREAM NAME	Residue to Steam & Power	Quench Water Blowdown	Catalyst Makeup
TOTAL FLOW RATE:			
Kg/hr	22905	69493	Confidential Information
BBL/SD			
TEMPERATURE °C	20	40	
PRESSURE kPa		330	
COMPONENT FLOW RATE:			
C_1			
C_2			
C_3			
C_4			
C_5 to 200°C			
200°C to 350°C			
350°C to 525°C	1366		
Solid hydrocarbon residue	14720		
Coal feedstock (maf)			
Unreacted coal (maf)	2722		
Ash	4097		
H_2			
H_2O		69477	
CO			
CO_2		3	
NH_3		13	
H_2S			
SO_2			
N_2			
Oxygen			

"H-COAL" - MATERIAL BALANCE

STREAM NO.	88	89	
STREAM NAME	Catalyst Disposal	Sulphur Plant Feed	
TOTAL FLOW RATE:			
Kg/hr	Confidential Information	38740	
BBL/SD			
TEMPERATURE °C			
PRESSURE kPa			
COMPONENT FLOW RATE:			
C_1		4083	
C_2		3478	
C_3		4280	
C_4		6462	
C_5 to 200°C		5266	
200°C to 350°C			
350°C to 525°C			
Solid hydrocarbon residue			
Coal feedstock (maf)			
Unreacted coal (maf)			
Ash			
H_2		457	
H_2O		2252	
CO		3498	
CO_2		7435	
NH_3			
H_2S		1529	
SO_2			
N_2			
Oxygen			

7.6 OPERATING REQUIREMENTS
H-COAL BASE CASE

7.6 BASE CASE OPERATING REQUIREMENTS

7.6.1 Utility Summary

7.6.1.1 Steam

Steam Produced (kg/hr)

Area		10 400 kPa	4 250 kPa	1 150 kPa	450 kPa
200	Coal Receiving & Preparation	-	-	-	-
300	Coal Liquefaction	35 446	-	-	172 800
400	Atmospheric & Vacuum Distillation	-	-	-	-
500	Naphtha Hydrotreating	-	-	-	-
600	Middle Distillate Hydrotreating	-	500	-	-
700	Heavy Distillate Hydrotreating	-	1 000	-	-
800	Light Ends Recovery	-	-	35 537[3]	-
900	Sulphur Plant	-	-	-	-
1000	Hydrogen Plant	17 420	-	45 690	-
1100	Oxygen Plant	-	101 106	36 058	44 069
1200	Steam & Power Generation	177 591	26 147[1]	204 310[1]	61 346[2]
1300	Phenol Recovery Plant	-	-	-	-
1400	Ammonia Recovery Plant	-	-	-	-
1500	Offsite Facilities	-	-	-	-
1600	Utilities	-	-	-	-
	TOTAL	230 457	128 753	321 595	278 215

Note:

1. The steam is produced by letting down 10 400 kPa from the high pressure power generation turbine.

2. This steam is produced by letting down 1 150 kPa from the medium pressure power generation turbine.

3. This steam is produced by flashing hot condensate.

Steam Consumed (kg/hr)

Area		10 400 kPa	4 250 kPa	1 150 kPa	450 kPa
200	Coal Receiving & Preparation	-	-	-	-
300	Coal Liquefaction	-	14 635	23 420	-
400	Atmospheric & Vacuum Distillation	-	-	25 300	-
500	Naphtha Hydrotreating	-	-	-	-
600	Middle Distillate Hydrotreating	-	-	290	-
700	Heavy Distillate Hydrotreating	-	-	400	-
800	Light Ends Recovery	-	114 118	12 160	-
900	Sulphur Plant	-	-	1 020	-
1000	Hydrogen Plant	-	-	-	55 575
1100	Oxygen Plant	-	-	-	-
1200	Steam & Power Generation	230 457	-	162 105	191 590
1300	Phenol Recovery Plant	-	-	-	-
1400	Ammonia Recovery Plant	-	-	96 900	-
1500	Offsite Facilities	-	-	-	20 700
1600	Utilities	-	-	-	10 350
	TOTAL	230 457	128 753	321 595	278 215

7.6.1.2 Water (Normal Operating Requirements)

10 400 kPa BFW, m^3/hr			Produced	Consumed
Area	300	Coal Liquefaction		39
	1000	Hydrogen Plant		19
	1200	Steam & Power Generation		197
	1600	Steam & Condensate Distribution	230	
	1600	BFW Plant	25	
			255	255
4 250 kPa BFW, m^3/hr				
Area	600	Middle Distillate Hydrotreating		0.6
	700	Heavy Distillate Hydrotreating		1.0
	1100	Oxygen Plant		112.4
	1600	Steam & Condensate Distribution	103	
	1600	BFW Plant	11	
			114	114
1 150 kPa BFW m^3/hr				
Area	1000	Hydrogen Plant		51
	1100	Oxygen Plant		40
	1600	Steam & Condensate Distribution	48	
	1600	BFW Plant	43	
			91	91
450 kPa BFW, m^3/hr				
Area	300	Coal Liquefaction		192
	1100	Oxygen Plant		49
	1600	Steam & Condensate Distribution	161	
	1600	BFW Plant	80	
			241	241

Demineralized Water, m³/hr			Produced	Consumed
Area	300	Coal Liquefaction	-	68
	1000	Hydrogen Plant	-	21
	1100	Oxygen Plant	-	42
	1200	Steam & Power Generation	-	19
	1600	BFW Plant	150	-
			150	150

Filtered Water, m³/hr

			Produced	Consumed
Area	1500	Cooling Water System	-	475
	1600	BFW Plant	-	178
	1600	Water Supply	653	-
			653	653

Raw Water m³/hr

			Produced	Consumed
Area	1600	Filtered Water System	-	653
	1600	Potable Water	-	15
	1600	Water Supply	668	-
			668	668

Cooling Water Distribution, m³/hr

			Produced	Consumed
Area	300	Coal Liquefaction	-	260
	400	Atmospheric & Vacuum Distillation	-	870
	500	Naphtha Hydrotreating	-	50
	600	Middle Distillate Hydrotreating	-	18
	800	Light Ends Recovery	-	2 907
	900	Sulphur Plant	-	90
	1000	Hydrogen Plant	-	4 165
	1100	Oxygen Plant	-	4 362
	1400	Ammonia Recovery	-	672
	1600	Utilities	-	194
	1600	Cooling Water	13 588	-
			13 588	13 588

7.6.1.3 Electrical Power (Normal Operating Requirements kW)

			Produced	Consumed
Area	200	Coal Receiving & Preparation	-	400
	300	Coal Liquefaction	-	21 500
	400	Atmospheric & Vacuum Distillation	-	3 000
	500	Naphtha Hydrotreater	-	1 600
	600	Middle Distillate Hydrotreating	-	4 400
	700	Heavy Distillate Hydrotreating	-	3 000
	800	Light Ends Recovery	-	14 600
	900	Sulphur Plant	-	1 800
	1000	Hydrogen Plant	-	29 400
	1100	Oxygen Plant	-	100
	1200	Steam & Power Generation	94 100	4 900
	1300	Phenol Recovery Plant	-	300
	1400	Ammonia Recovery Plant	-	700
	1500	Offsites	-	3 800
	1600	Utilities		4 600
			94 100	94 100

7.6.1.4 Fuel Gas (Normal Operating Requirements 10^3 kJ/hr)

			Produced	Consumed
Area	300	Coal Liquefaction	-	220 000
	400	Atmospheric & Vacuum Distillation	-	275 000
	500	Naphtha Hydrotreater	-	30 300
	600	Middle Distillate Hydrotreating	-	60 600
	700	Heavy Distillate Hydrotreating	-	42 900
	900	Sulphur Plant	-	6 100
	1100	Oxygen Plant	-	490 000
	1300	Phenol Recovery Plant	15 800	-
	1500	Offsites - Flares	-	10 000
	1600	Fuel Gas System	1 821 200	-
			1 837 000	1 134 900
		excess fuel gas		702 100

7.6.2 Catalysts and Chemicals

Area	Item	Description	Annual Operating Requirements
300	Coal Liquefaction Catalyst	American Cyanamid HDS-1442A catalyst	4 995 000 kg
	Acid Gas Absorbent	DGA (diglycolamine)	4 200 kg
500	Naphtha Hydrotreater catalyst	Proprietary Information	$ 282 700[1]
600	Middle Distillate Hydrotreater catalyst	Proprietary Information	$ 277 500[1]
700	Heavy Distillate Hydrotreater catalyst	Proprietary Information	$ 143 300[1]
800	PSA Adsorbent	Proprietary Information	$ 56 400[2]
	Molecular Sieve Adsorbent	Proprietary Information	$ 17 200[2]
900	Stretford Plant Chemicals	Proprietary Information	48 150 kg
1000	Sour Gas Shift Catalyst	Proprietary Information	70 160 kg
	Sour Gas Solvent	Proprietary Information	55 700 kg
1200	FGD Neutralizer	Lime	$366 000
1500 & 1600	Ion Exchange Regenerator and Cooling Water pH Control Agent	93% sulphuric acid	1 379 000 kg
	Water Softener	Lime	1 902 000 kg
	Ion Exchange Regenerator	100% Caustic Soda	94 300 kg
	Oxygen Scavenger	Hydrazine	211 460 L
	BFW Chemicals	Corrosion & fouling inhibitors	$ 19 900
	Biocide	Betz C-30	1 680 kg
	Disinfectant	Sodium Hypochlorite 15%	14 300 kg
	Nutrient for Biological Effluent Treatment	Phosphoric acid	6 520 kg
	Algaecide	Chlorine	4 490 kg

Area	Item	Description	Annual Operati Requirements
	Demineralizer Material	Cation ion exchange resin	440 m^3 [3]
	Demineralizer Material	Anion ion exchange resin	290 m^3 [3]

NOTES:
1. Based on 3 years expected life.
2. Based on 5 years expected life.
3. Based on 7 years expected life.

7.6.3 Plant Personnel Requirements

7.6.3.1 Office Staff & Day Personnel

Management	10
Technical Staff	73
Office Staff	45
Laboratory Personnel	17
Maintenance Personnel	300
TOTAL:	445

7.6.3.2 Shift Operators & Maintenance Personnel

Process Units	200
Utilities	75
Offsites	25
Maintenance	75
Laboratory	15
Miscellaneous (firefighting, security, etc.)	25
TOTAL:	415

TOTAL PLANT PERSONNEL	860

7.7 OPTIONAL PROCESS CONFIGURATIONS

The different process options investigated are briefly described in this section. Although, a detailed description of the Base Case H-Coal process has been given in Sections 7.2 to 7.6, a brief description of the key features relating to energy utilization is presented in Section 7.7.1.

The subsequent sections briefly describe the parameters varied in the optional cases. Overall material balances have been developed and presented in simplified block flow diagrams.

The key features of the different options are listed in Tables 7.1 and 7.2.

7.7.1 H-Coal Base Case Configuration

The Base Case configuration of the H-Coal liquefaction plant is designed to be energy self sufficient. While no energy is imported, 40% distillate yield on maf coal basis is obtained.

With the coal feed as the only form of energy, the residue and fuel gas produced in the liquefaction process is used in the complex to fulfill the energy needs.

With the relatively low distillate yield, substantial quantity of residue is formed. Most of the residue is gasified to produce hydrogen. A portion of this hydrogen is used in the liquefaction stage, and the remainder is utilized in upgrading the products.

In addition to hydrogen production, power as required, is generated by combustion of the residue. The excess residue is disposed of in an environmentally acceptable manner.

The liquefaction plant also produces significant quantities of fuel gas. Besides the fired heaters and other uses, the fuel gas is used to drive the gas turbines in the air separation plant - a very large gas consumer. The exhaust gases from the air separation compressor turbine drivers possess a high energy value, which is used to generate make-up steam. The excess fuel gas is exported from the liquefaction complex.

Drawing No. 100-43-001 (Page 7-44) is a detailed block flow diagram for the H-Coal Base Case configuration. The detailed description of the process is given in Section 7.3 of this report.

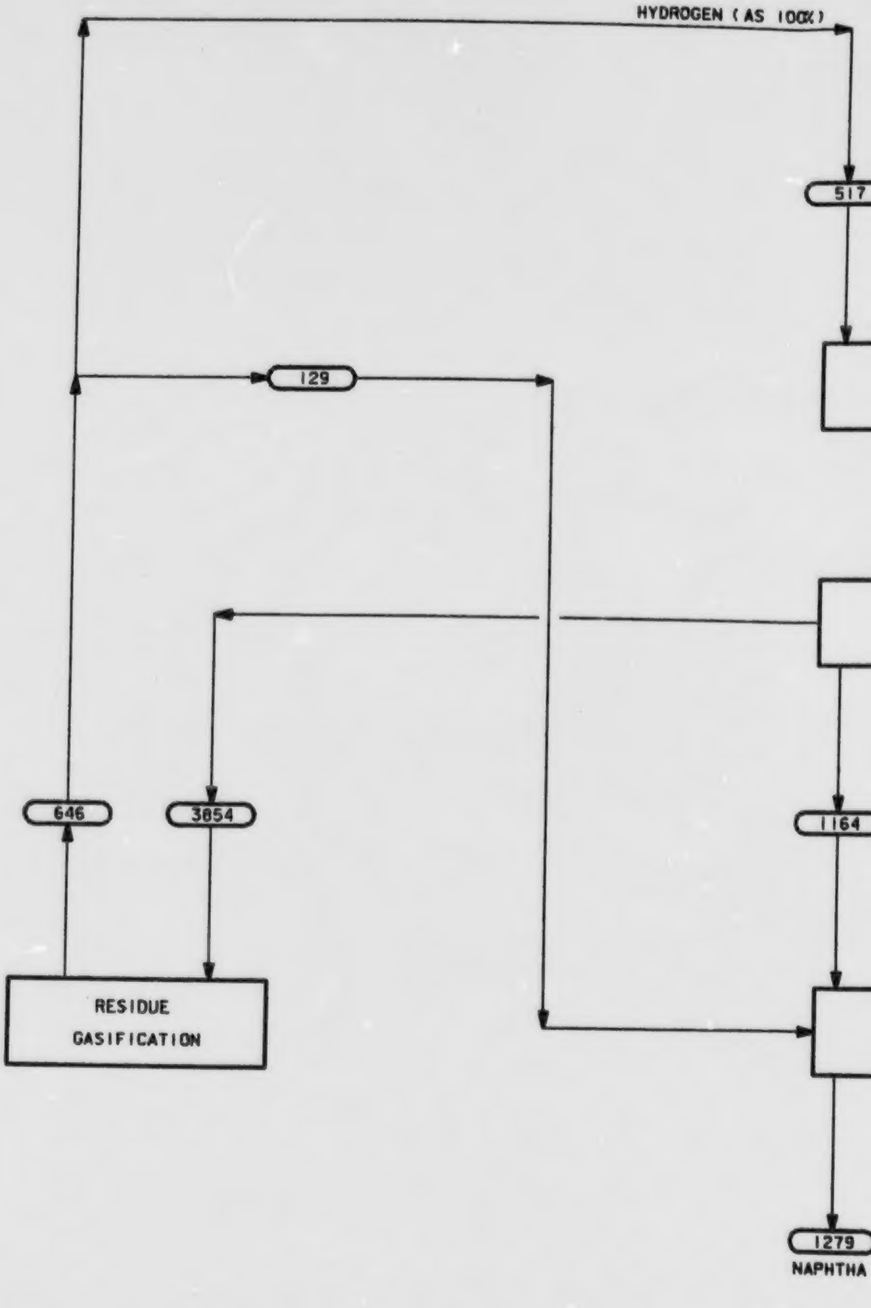
7.7.2 Case 1 - H-Coal Configuration with 50% Distillate Yield

With the increased distillate yield, a lesser amount of residue is produced. With the increased distillate yield the hydrogen requirement for both the coal conversion and product upgrading is also increased.

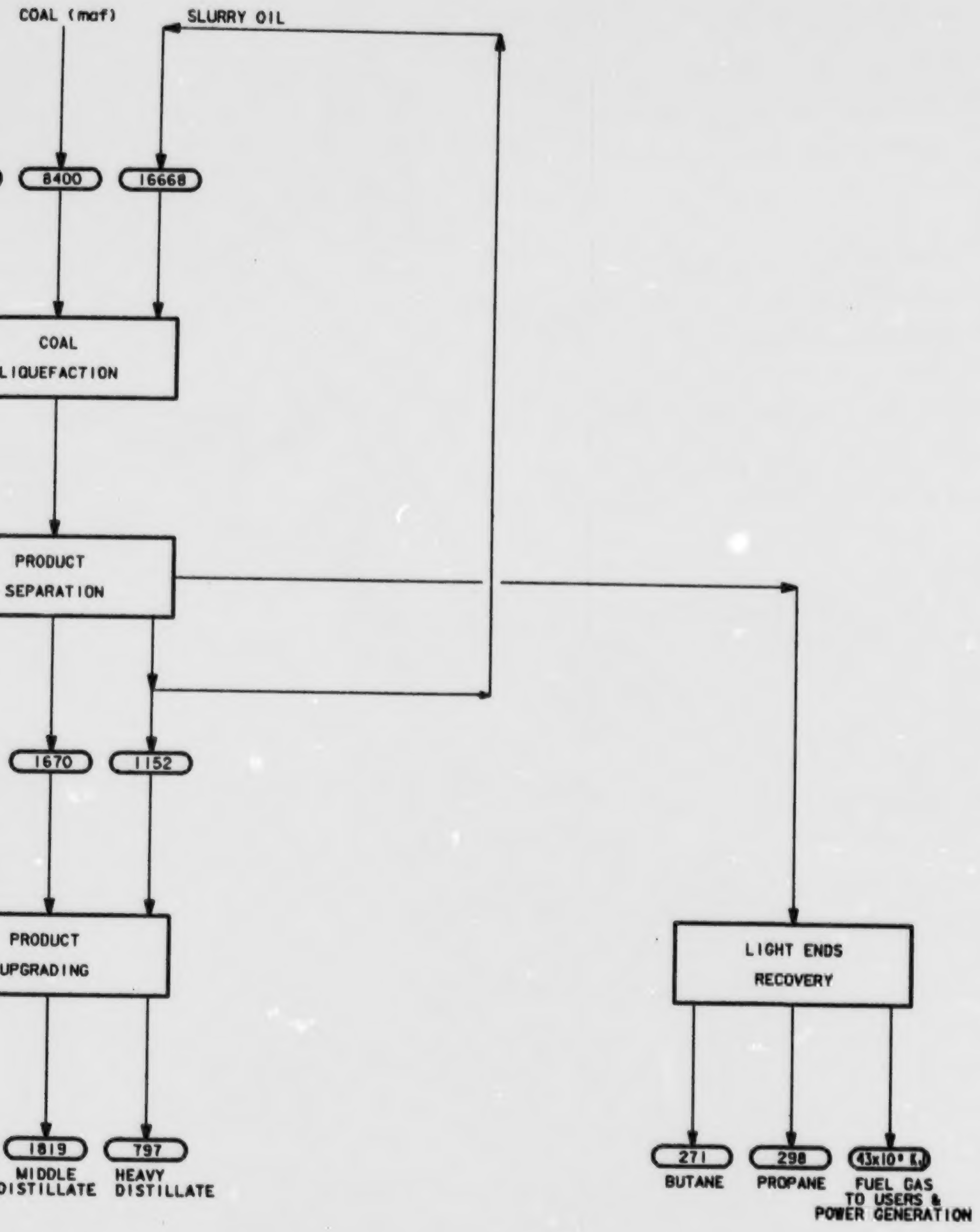
Gasification of the residue produces all the hydrogen needs for the plant. In the Base Case, power was generated by combustion of the residue (beyond the hydrogen needs), and the fuel gas was used to fulfill the oxygen plant needs while exporting the excess. In this process option, the oxygen plants needs and the incidental process requirements are fulfilled by the fuel gas. Sufficient fuel gas is available to generate power for the plant requirements.

Hence, this option is energy self sufficient.

Drawing No. 100-43-005 shows the simplified block flow diagram plus an overall material balance.


HYDROGEN (AS 100%)
517
129
646
3854
1164
RESIDUE GASIFICATION
1279
NAPHTHA

UNLESS SPECIFIED OTHERWISE, ALL NUMBERS REPRESENT TONNE PER DAY.


COAL (maf)
SLURRY OIL
8400
16668
COAL
LIQUEFACTION
PRODUCT
SEPARATION
1670
1152
PRODUCT
UPGRADING
1819
MIDDLE
DISTILLATE
797
HEAVY
DISTILLATE
LIGHT ENDS
RECOVERY
271
BUTANE
298
PROPANE
43×10^{6} KJ
FUEL GAS
TO USERS &
POWER GENERATION

Western Canada Coal Liquefaction Study
Prefeasibility Study
Simplified Block Flow Diagram
Case I - H-Coal process With
50% Distillate Yield
Drawing No. 100-43-005

7.7.3 Case 2 - 40% Distillate Yield, Residue Gasification and Partial Oxidation of Fuel Gas

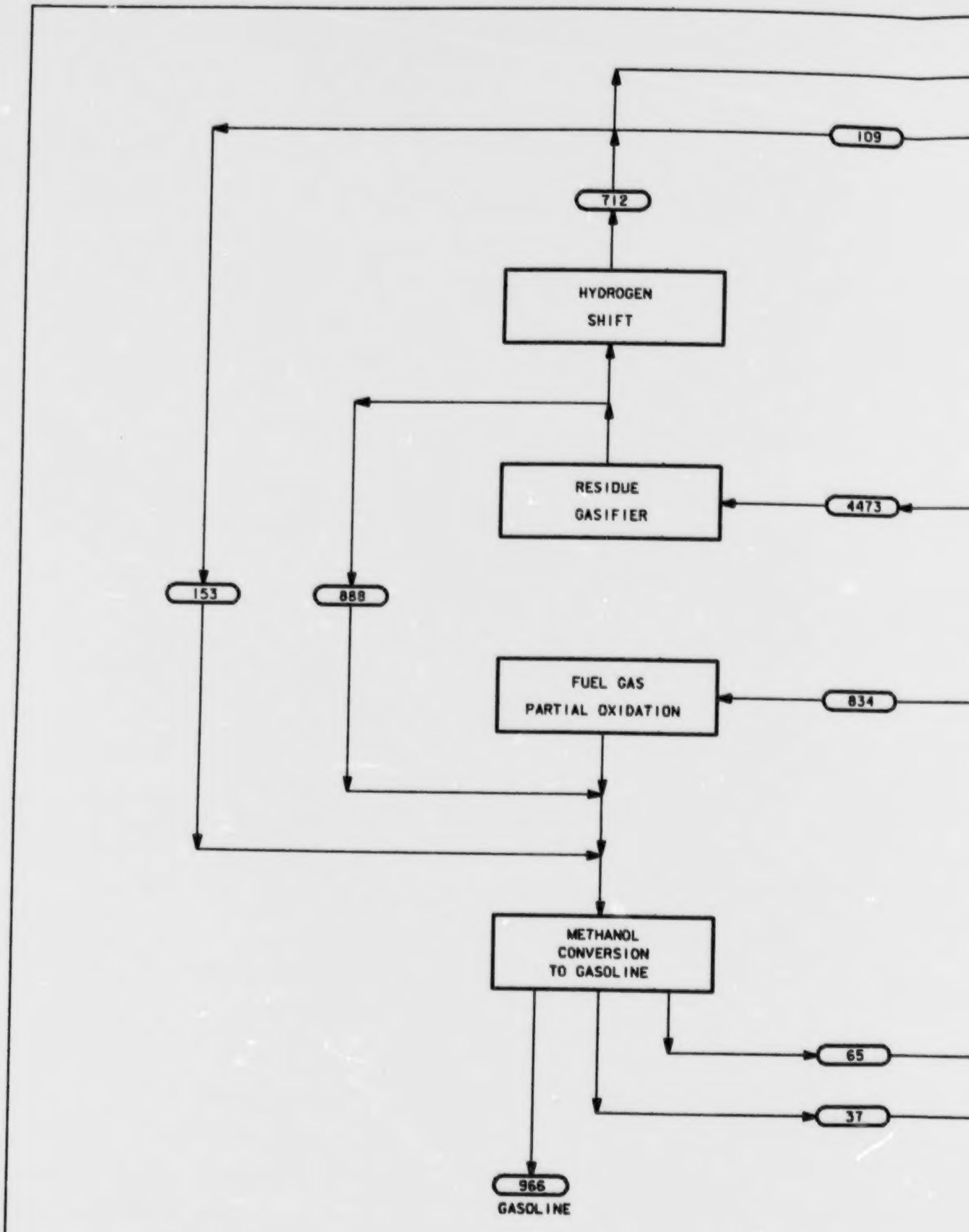
In this option, all the residue produced is gasified to form a synthesis gas and finally hydrogen. All the fuel gas produced in the liquefaction process is also converted to syngas via partial oxidation.

The syngas from the gasified residue contains H_2 and CO in a 1:1 ratio. A slip stream of this gas, after a small dedicated acid gas removal unit, is fed to the Mobil MTG complex, as described below. The bulk sour gas stream undergoes shifting for maximum H_2 production. After sweetening and purification, high purity hydrogen is utilized in the liquefaction unit and the product upgrading facility. Excess hydrogen is fed to the MTG complex to adjust the H_2/CO ratio.

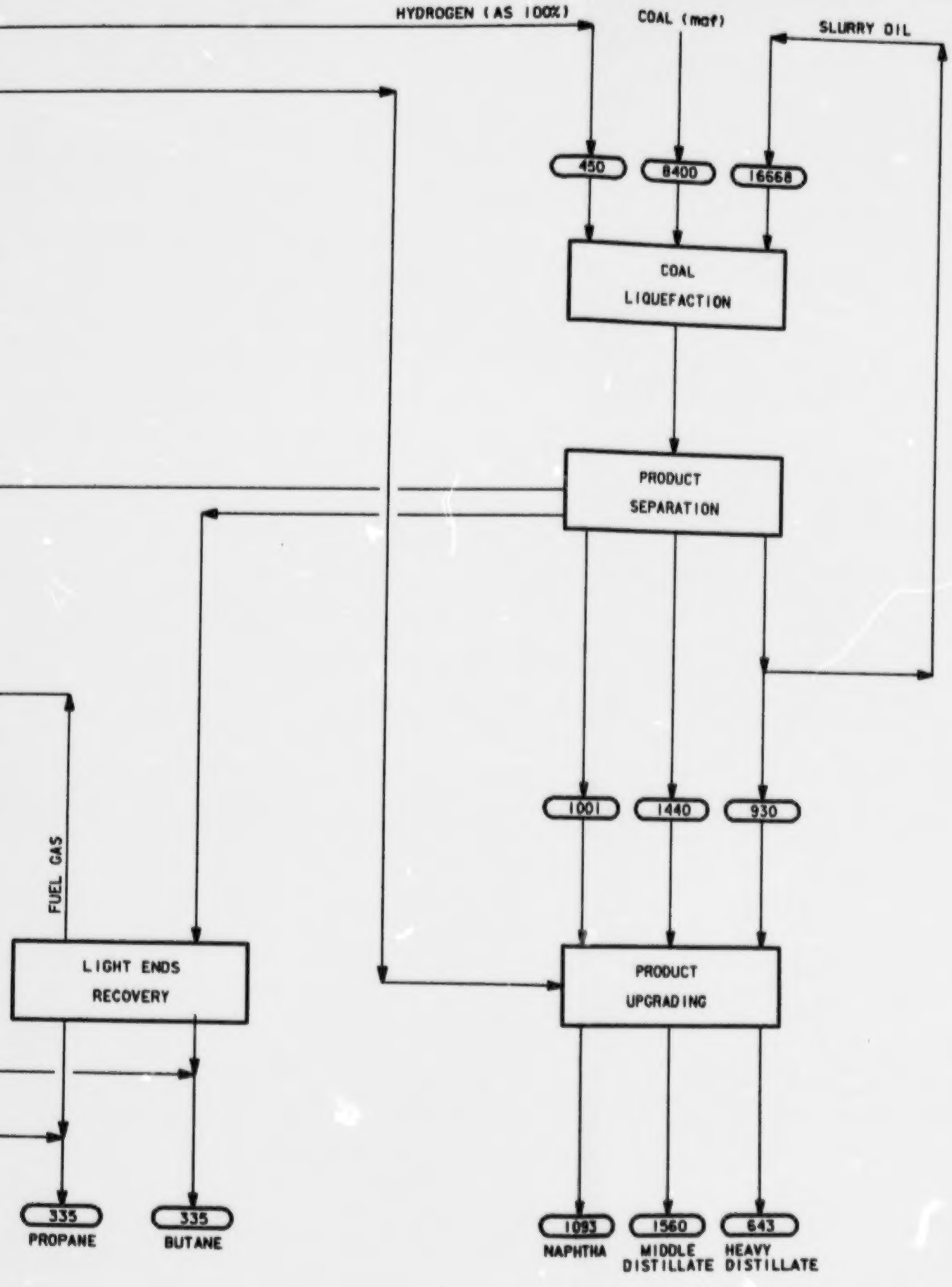
All the fuel gas produced in the liquefaction process and the MTG plant is fed to a partial oxidation unit. Here the hydrocarbon gases react with the oxygen and steam and are partially oxidized to produce a syngas in the H_2/CO ratio of about 1.5:1. This syngas is mixed with the high purity hydrogen stream and the sweetened slip stream from the residue gasification in an appropriate proportion to obtain a 2.2:1 H_2 to CO ratio as feed to the MTG complex. In the Mobil MTG process, syngas is converted to gasoline blendstock via the route syngas-methanol-gasoline, using Mobil proprietary technology.

Since all the residue is gasified, electricity has to be imported to meet the needs of the plant. Also the compressors in the air separation plant have to be driven with imported power. However, the steam requirement within the complex is generated from the partial oxidation plant. The combustion gases from the partial oxidation plant do not contain any solids, therefore, do not require quenching. The gas contains sufficient heat to produce make-up steam for the complex.

A simplified block flow diagram with an overall material balance is shown in Drawing No. 100-43-006.


109
712
HYDROGEN SHIFT
RESIDUE GASIFIER
4473
153
888
FUEL GAS PARTIAL OXIDATION
834
METHANOL CONVERSION TO GASOLINE
65
37
966
GASOLINE

ALL NUMBERS REPRESENT TONNE PER DAY.


HYDROGEN (AS 100%)
COAL (maf)
SLURRY OIL
450
8400
16668
COAL LIQUEFACTION
PRODUCT SEPARATION
FUEL GAS
1001
1440
930
LIGHT ENDS RECOVERY
PRODUCT UPGRADING
335
PROPANE
335
BUTANE
1093
NAPHTHA
1560
MIDDLE DISTILLATE
643
HEAVY DISTILLATE

Western Canada Coal Liquefaction Study
Prefeasibility Study
Simplified Block Flow Diagram
Case 2 - H-Coal Plus MTG Configuration
Drawing No. 100-43-006

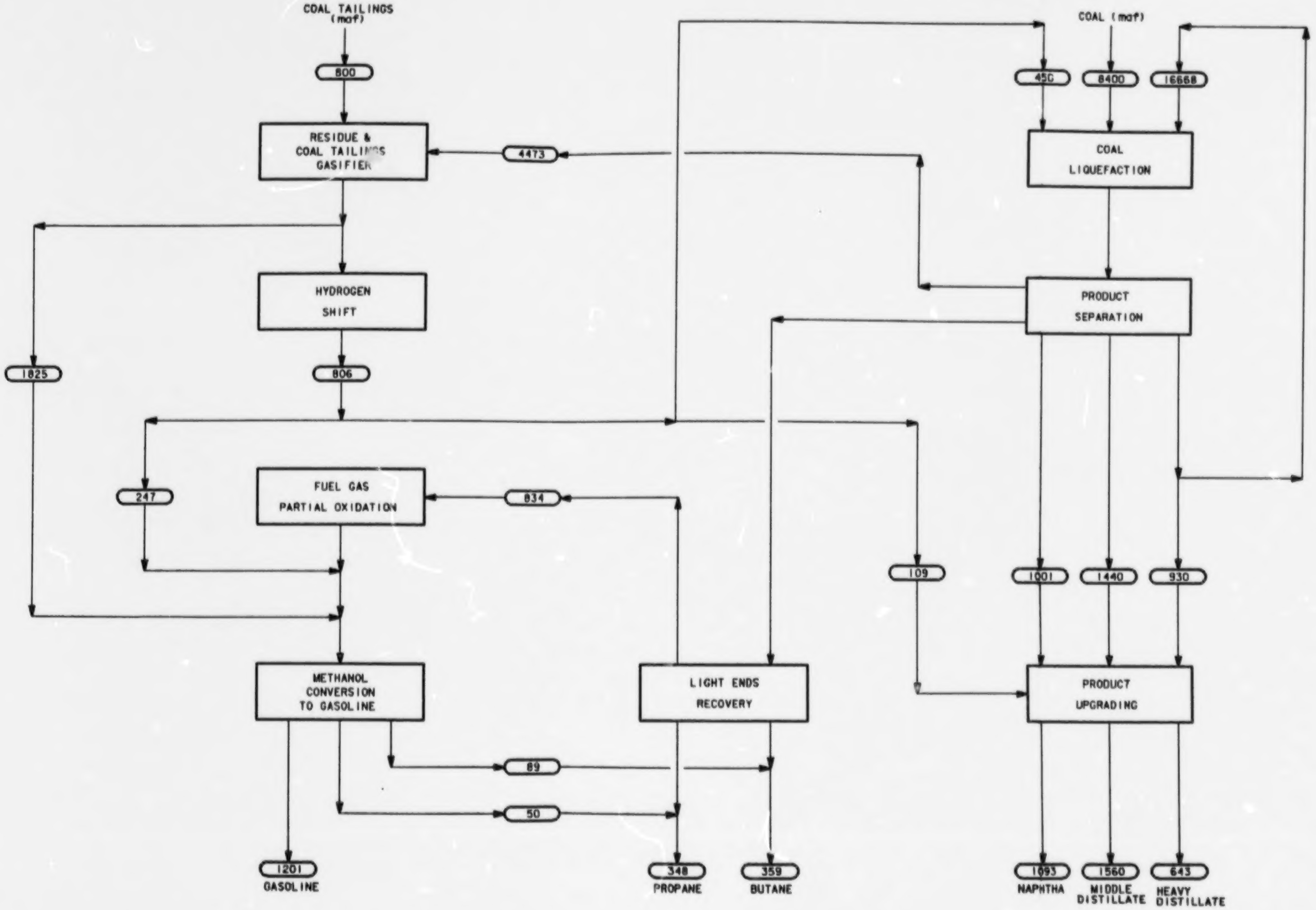
7.7.4 Case 3 - 40% Distillate Yield, Gasification of Residue and Coal Beneficiation Rejects and Partial Oxidation of Fuel Gas

This process option is similar to Case 2 except that, along with the residue, the tailings from the coal beneficiation plant are included for gasification. Hence, the gasifier capacity and throughput is increased.

The material balance around the syngas is adjusted to obtain a 2.2:1 H_2/CO ratio prior to feeding the Mobil MTG facility as described in Section 7.7.3

This option increases the "feed" to the syngas production and hence increases the blended gasoline product from the MTG plant. In this option, power is imported to compensate for the residue and fuel gas utilization for increased product yields.

A simplified block flow diagram with an overall material balance is shown in Drawing No. 100-43-007.


COAL TAILINGS (maf)
800
RESIDUE & COAL TAILINGS GASIFIER
4473
COAL (maf)
450
8400
16668
COAL LIQUEFACTION
PRODUCT SEPARATION
HYDROGEN SHIFT
1825
806
247
FUEL GAS PARTIAL OXIDATION
834
109
1001
1440
930
METHANOL CONVERSION TO GASOLINE
LIGHT ENDS RECOVERY
PRODUCT UPGRADING
89
50
1201
GASOLINE
348
PROPANE
359
BUTANE
1093
NAPHTHA
1560
MIDDLE DISTILLATE
643
HEAVY DISTILLATE

ALL NUMBERS REPRESENT TONNE PER DAY.

Western Canada Coal Liquefaction Study
Prefeasibility Study
Simplified Block Flow Diagram
Case 3 - H-Coal Plus MTG Configuration
With Coal Rejects
Drawing No. 100-43-007

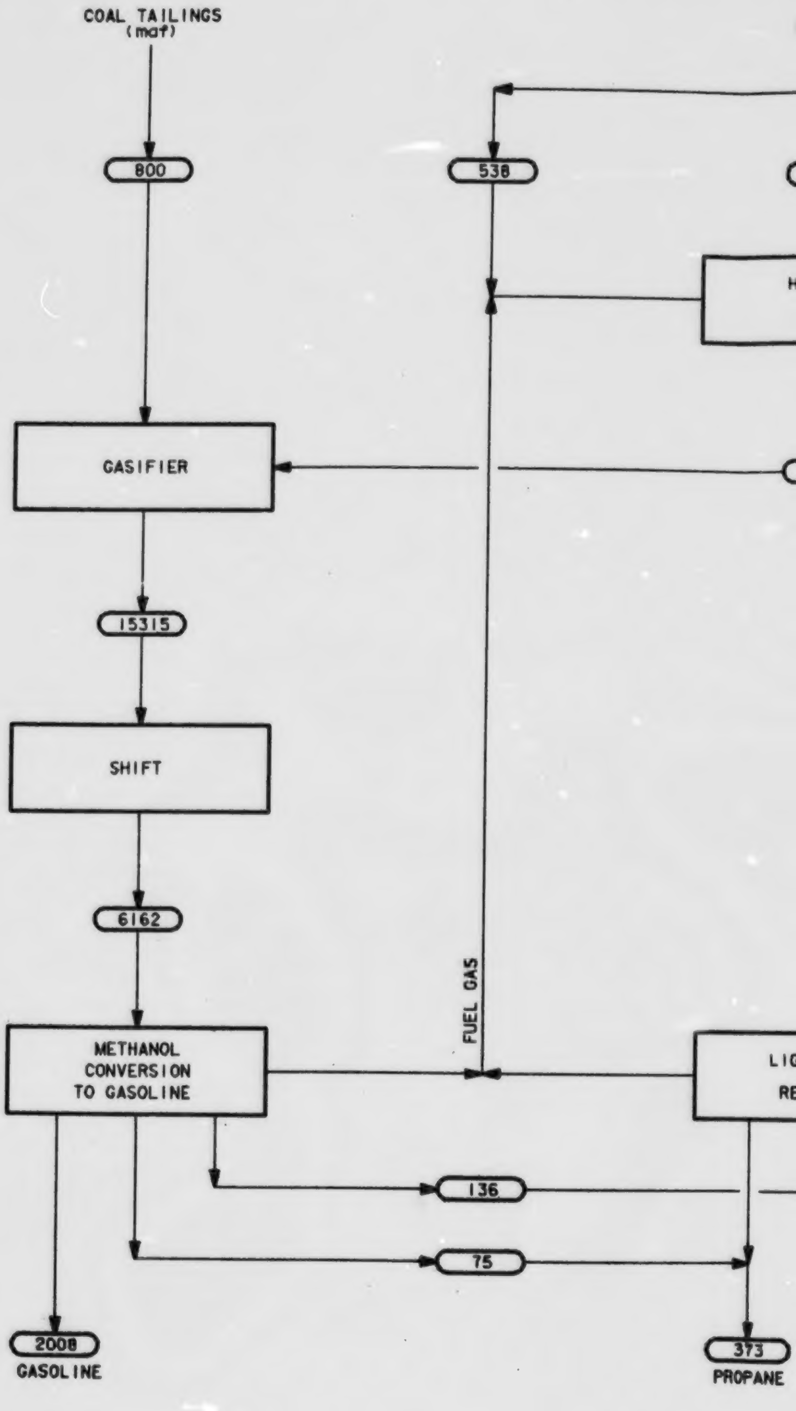
7.7.5 Case 4 - 40% Distillate Yield, Hydrogen Production via Steam Reforming of Natural Gas and Gasification of Residue and Coal Beneficiation Rejects for Production of Gasoline Blendstock

In this process option, all the residue from the liquefaction process and the coal beneficiation rejects is gasified and shifted to a 2.2:1 H_2/CO ratio. After acid gas removal, the syngas is fed to a Mobil MTG plant for the production of gasoline blendstock.

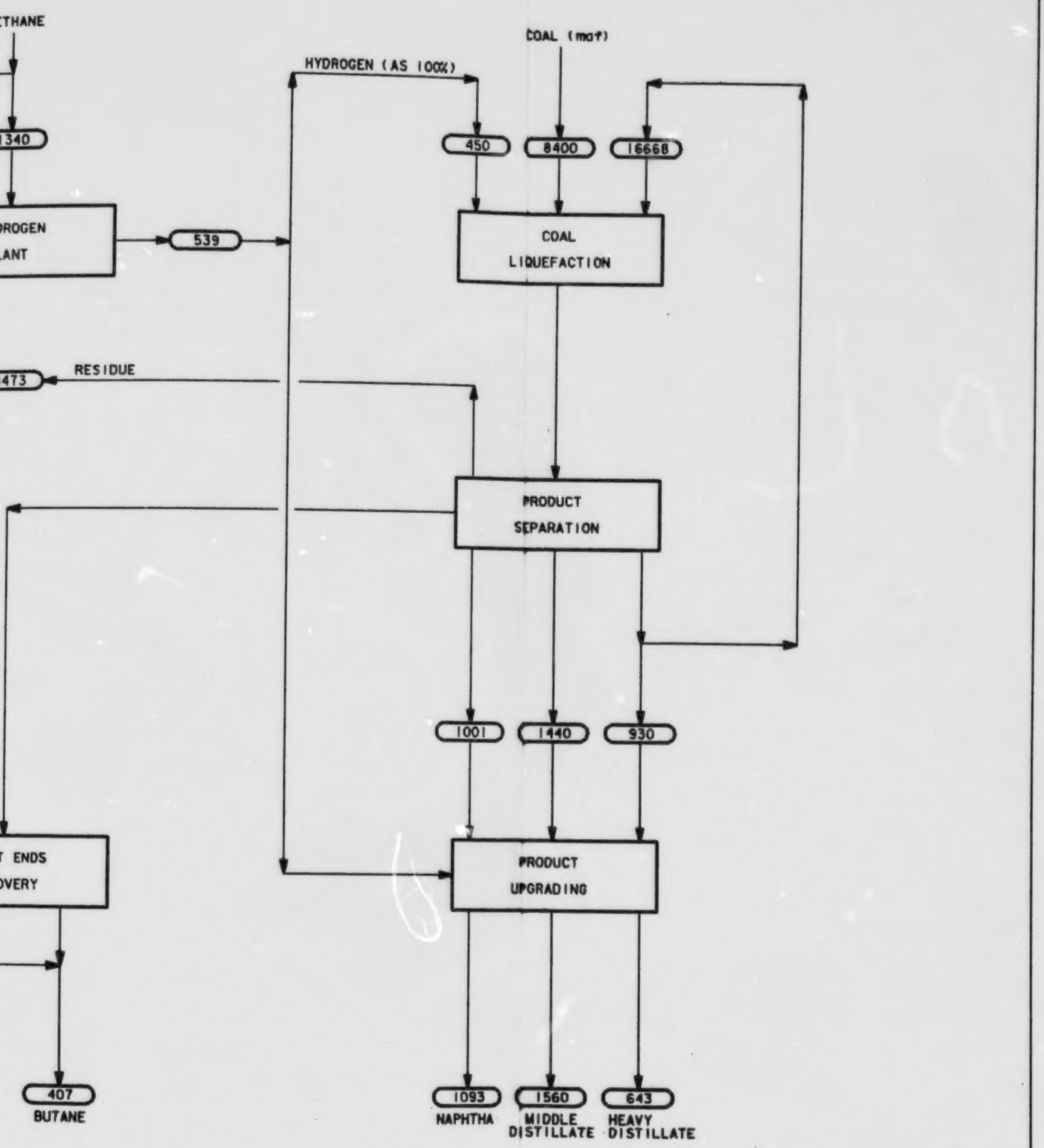
The major difference in this configuration is the hydrogen production plant. Hydrogen for the coal conversion unit and the hydrogen requirement for the upgrading of the liquid products is produced by reforming imported natural gas. The fuel gas requirements for the steam reformer is partially met by the fuel gas produced in the liquefaction and the MTG plant. The deficiency is met by imported natural gas.

The power requirement for the liquefaction complex is met by importing electricity.

A simplified block flow diagram with a general material balance is shown in Drawing No. 100-43-008.


COAL TAILINGS (maf)
800
538
H
GASIFIER
15315
SHIFT
6162
METHANOL CONVERSION TO GASOLINE
FUEL GAS
LIG
RE
136
75
2008
GASOLINE
373
PROPANE

ALL NUMBERS REPRESENT TONNE PER DAY.


ETHANE
1340
HYDROGEN (AS 100%)
COAL (maf)
450
8400
16668
DROGEN
LANT
539
COAL
LIQUEFACTION
473
RESIDUE
PRODUCT
SEPARATION
1001
1440
930
T ENDS
OVERY
PRODUCT
UPGRADING
407
BUTANE
1093
NAPHTHA
1560
MIDDLE
DISTILLATE
643
HEAVY
DISTILLATE

Western Canada Coal Liquefaction Study
Prefeasibility Study
Simplified Block Flow Diagram
Case 4 - H - Coal Plus MTG Configuration
With H_2 Production From Natural Gas
Drawing No. 100-43-008

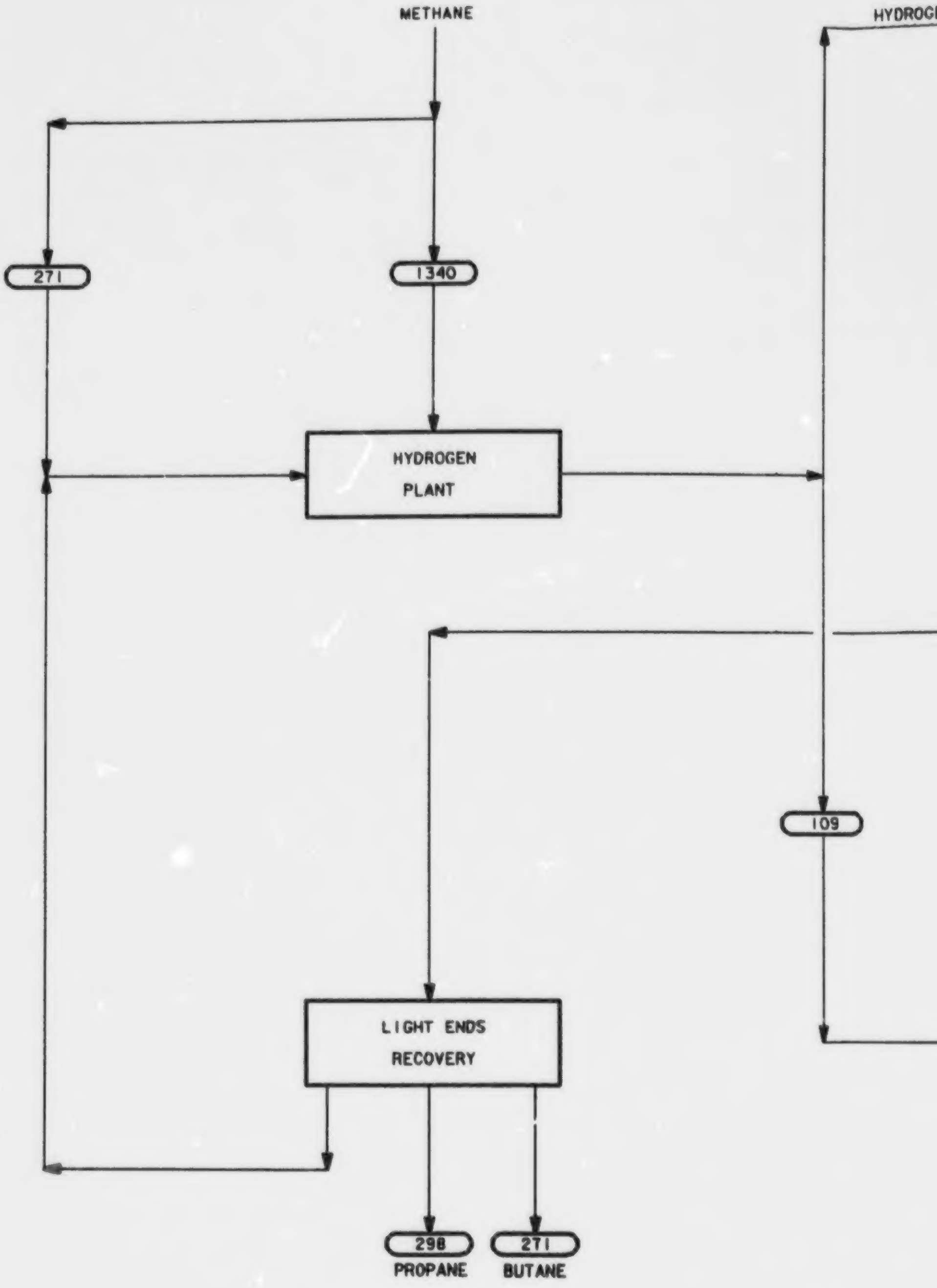
7.7.6 Case 5 - 40% Distillate Yield and Hydrogen Production via Steam Reforming of Natural Gas

In this option, hydrogen is produced by steam reforming of imported natural gas. The fuel requirements for the steam reformer is partially met by the fuel gas produced by the liquefaction plant. The difference is met by imported natural gas.

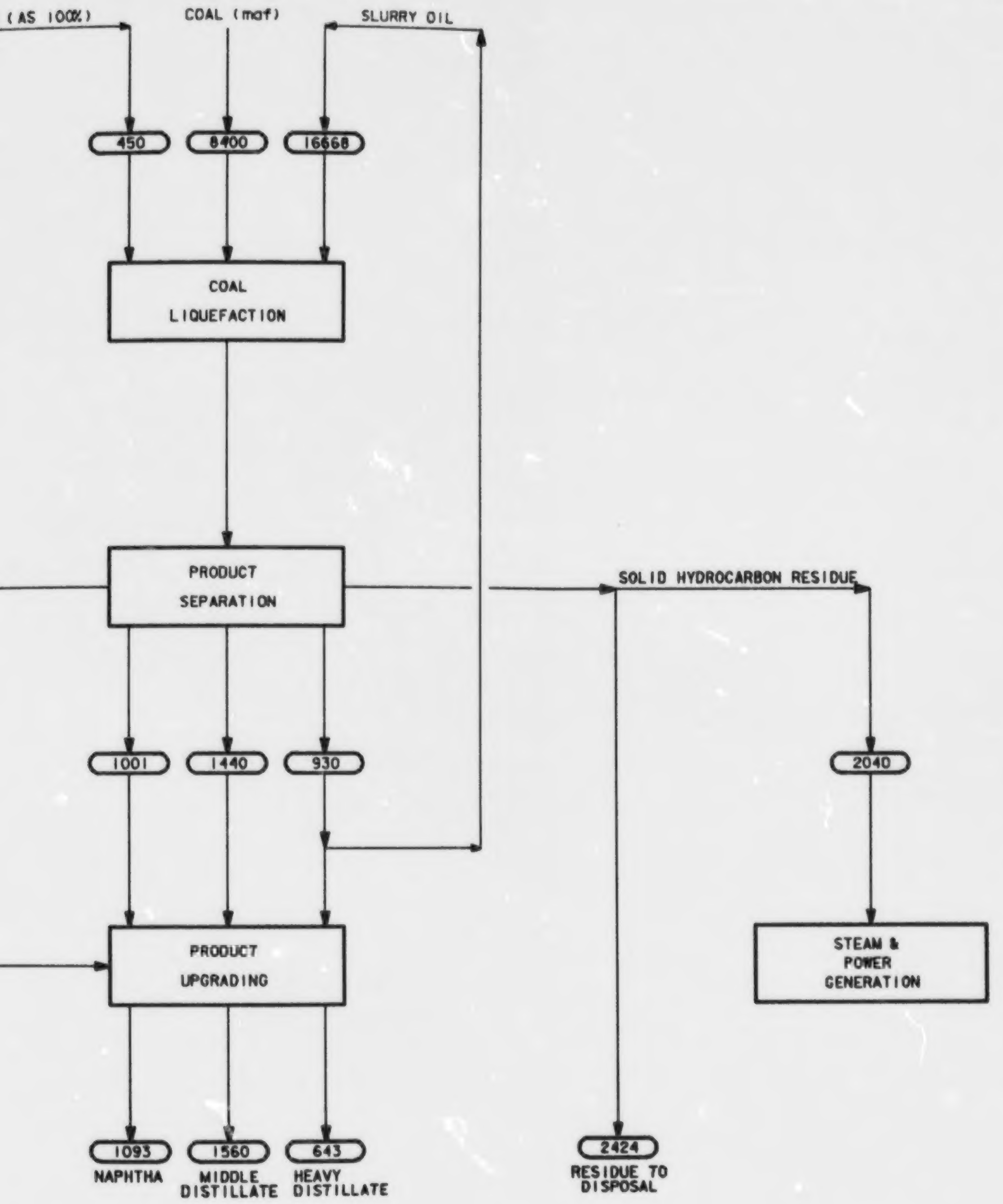
The fuel requirement for the fired heaters and other users in the liquefaction plant is fulfilled by part of the fuel gas produced within the complex. The excess fuel gas is used in the steam reformer for hydrogen production, as noted above.

Steam and power for the plant needs are generated by combusting part of the residue from the coal liquefaction process. The excess residue is disposed of at zero value or zero disposal cost. (It is assumed that an environmentally acceptable use can be found for this material under these conditions -- i.e. power generation excess to the needs of the liquefaction facility).

A simplified block flow diagram with a general material balance is shown in Drawing No. 100-43-009.


METHANE
HYDROG
271
1340
HYDROGEN PLANT
109
LIGHT ENDS RECOVERY
298
PROPANE
271
BUTANE

ALL NUMBERS REPRESENT TONNE PER DAY.


(AS 100%)
COAL (maf)
SLURRY OIL
450
8400
16668
COAL LIQUEFACTION
PRODUCT SEPARATION
SOLID HYDROCARBON RESIDUE
1001
1440
930
2040
PRODUCT UPGRADING
STEAM & POWER GENERATION
1093
NAPHTHA
1560
MIDDLE DISTILLATE
643
HEAVY DISTILLATE
2424
RESIDUE TO DISPOSAL

Western Canada Coal Liquefaction Study
Prefeasibility Study
Simplified Block Flow Diagram
Case 5 - H Coal Process With H_2 Production
From Natural Coal
Drawing No. 100-43-009

TABLE 7.1

PROCESS OPTION CONFIGURATION

	Base Case	Case 1	Case 2	Case 3	Case 4	Case 5
Residue	To H_2 Production To Power Plant To Disposal *	To H_2 Production	To H_2 Production To Methanol Conversion	To H_2 Production To Methanol Conversion	To Methanol Conversion To Disposal *	To Power Plant
Coal Rejects	-	-	-	To H_2 Production To Methanol Conversion	To Methanol Conversion	-
Excess Fuel Gas	To O_2 Plant To Export	To O_2 Plant To Power Plant	To Partial Oxidation -	To Partial Oxidation -	To Steam Reformer as Fuel Gas To O_2 Plant	To Steam Reformer as Fuel Gas
Natural Gas	-	-	-	-	To H_2 Production To Fuel Gas Make-up	To H_2 Production To Fuel Gas Make-up
Power	From Power Plant	From Power Plant	Import	Import	Import	From Power Plant

* Residue disposal in an environmentally acceptable manner.

TABLE 7.2

RAW MATERIALS AND PRODUCTS

	Base Case	Case 1	Case 2	Case 3	Case 4	Case 5
Raw Materials						
Coal MAF (T/D)	8 400	8 400	8 400	8 400	8 400	8 400
Tailings MAF (T/D)	-	-	-	800	800	-
Natural Gas (KJ x 10^6 per day)	-	-	-	-	69 784	69 784
Utilities (Import)						
Power (MW)	-	-	237	307	155	-
Fuel Gas (KJ x 10^6 per day)	-	-	-	-	28 018	14 113
Products (T/D)						
Naphtha	1 093	1 279	1 093	1 093	1 093	1 093
Middle Distillate	1 560	1 819	1 560	1 560	1 560	1 560
Heavy Distillate	643	797	643	643	643	643
Propane	298	298	335	348	373	298
Butane	271	271	335	359	407	271
Gasoline Blendstock	-	-	966	1 201	2 008	-
Fuel Gas (KJ x 10^6 per day)	17 160	-	-	-	-	-

Infrastructure and Support Facilities

8.0 INFRASTRUCTURE AND SUPPORT FACILITIES

8.1 GENERAL

As Greenhills coal was chosen as the study reference coal a site adjacent to the Greenhills mine in south-eastern British Columbia was selected for the coal liquefaction evaluation. As a result, infrastructure described in this section is specific to this location.

The support facilities relating to the Base Case H-Coal evaluation include coal receiving and storage, steam and power generation and all necessary offsite and utility facilities to service the Process Units. These facilities include pollution control systems such as flue gas desulphurization, effluent water treatment, and process flare systems; and service systems, such as catalyst and chemical storage, product tankage and loadout, steam and condensate distribution, cooling water, BFW and water treatment systems.

8.2 INFRASTRUCTURE SITE FACTORS

Land is currently available in the area of the Greenhills mine for development of the coal liquefaction complex. Figure 8.1 shows the Greenhills area, the developing mine, potential sites for the coal liquefaction complex and other facilities. As property held by B.C. Coal is zoned at present for coal development, no problem is anticipated in rezoning a section of the land within the B.C. Coal property for the coal liquefaction facility. This is an important factor to consider in selecting a site since it can substantially shorten the time required to obtain the necessary permits. Estimated size of the site is 120 hectares, consisting of 30 hectares for process units, 20 hectares for effluent and fire water ponds and 70 hectares for coal receiving, offsites and utilities. Depending on site clearance requirements, facilities such as effluent ponds and product loadout can be located remote from the process unit site.

One of the primary reasons for considering a site adjacent to the mine is that beneficiated coal can be delivered by belt conveyor avoiding the cost of rail shipment. Current rail service to the Sparwood coal reserve is provided by a branch line off the Canadian Pacific mainline located 20 km south of the complex. A spur off the branch line intended to load coal from the Greenhills mine could be extended to provide product loadout facilities for the coal liquefaction complex. The suitability of this rail line however for shipment of large tonnage equipment for the complex will have to be evaluated. All-weather road access is also available to the plant site for transport of personnel and equipment during the construction phase; personnel, raw materials and product in the operating phase.

Raw water for the complex is available in part (270 cubic metres per hour) from wells on the mine property and the remainder (400 cubic metres per hour) from the Elk River. Water quality is typical of ground water with moderate concentrations of limestone.

Power available in the Greenhills area of the Elk River Valley is limited and supplied by a 230 kV line from the Calgary Power System on an exchange agreement with B.C. Hydro. In the Base Case emergency power would be drawn from this line for non-production purposes. Should a larger power supply be required, as was investigated in some of the alternative cases, a new 230 kV line, would be provided from a substation southeast of Natal B.C., 40 km from the complex.

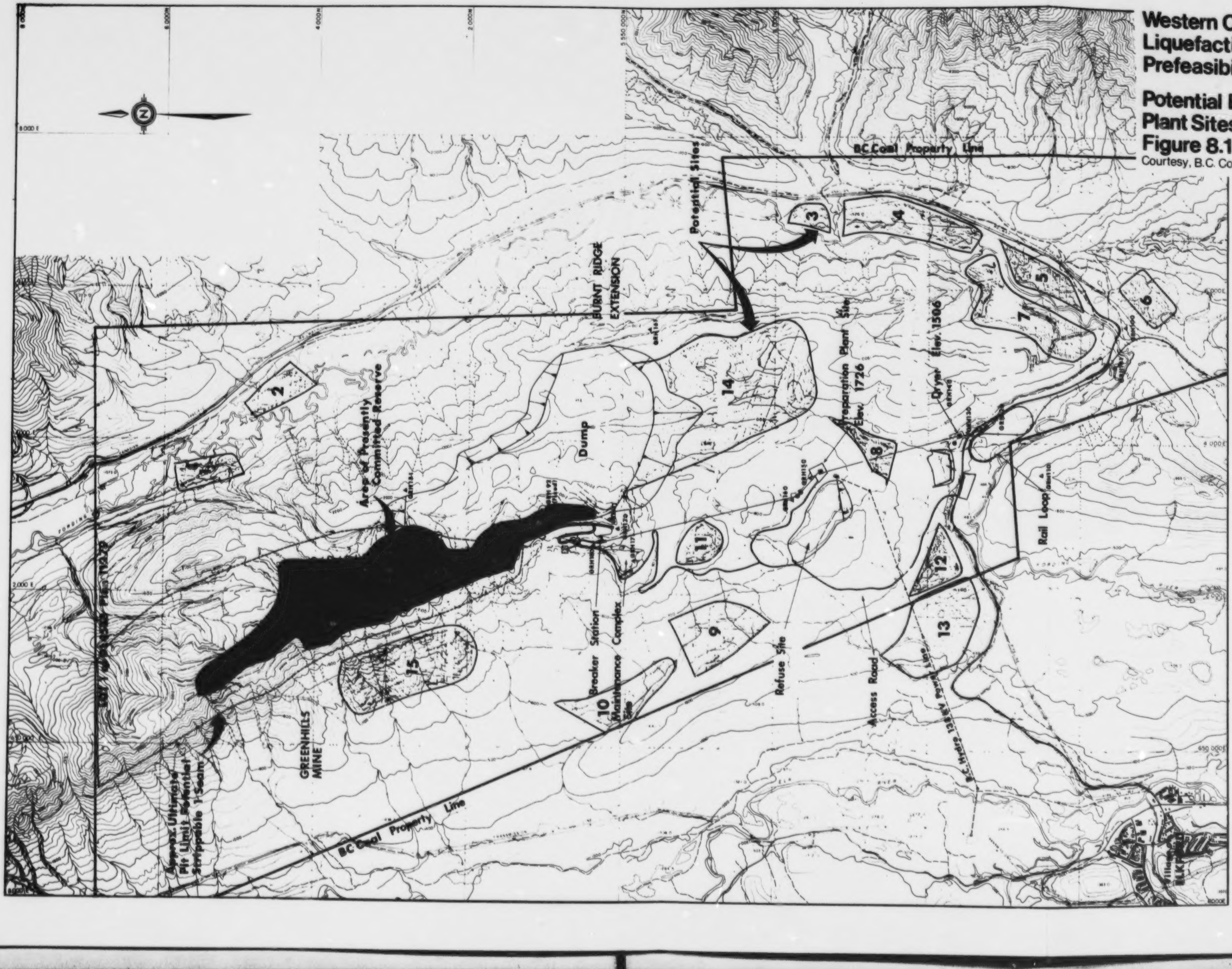
Excess fuel gas produced in the Base Case will be transported from the complex by pipeline for consumption locally at the mine or nearby industry. Should natural gas be required by the complex as indicated in some alternate configurations, a pipeline would be provided from the Columbia Natural Gas pipeline located 20 km south of the proposed site.

Western Canada Coal Liquefaction Study Prefeasibility Study

Potential Liquefaction Plant Sites

Figure 8.1

Courtesy, B.C. Coal


BC Coal Property Line
Potential Sites
BURNT RIDGE EXTENSION
Preparation Plant Site Elev. 1726
Dryer Elev. 1506
Area of Presently Committed Reserve
Dump
Rail Loop
Breaker Station
Maintenance Complex Site
Refuse Site
Access Road
B.C. Hydro 138 KV Power Line
GREENHILLS MINE
Amax Ultimate Pit Limit Potential Strippable 1-Seam
BC Coal Property Line
Village of ELKFORD
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8.3 SUPPORT FACILITIES

8.3.1 Area 200 Coal Receiving & Handling

8.3.1.1 Coal Handling

Beneficiated coal is received from the adjacent coal mine by belt conveyor at a design rate of 500 tonne per hour. Coal feed rate to the coal liquefaction unit is 417 tonne per hour with a design ash content of eight weight percent, and a design moisture content of eight weight percent. Size range of the received coal is minus 50 mm. The design delivery rate of 500 tonne per hour is specified to permit operating flexibility and allow for downtime of mine equipment.

Upon receipt at the plant, the coal is weighed automatically and transferred either directly to one of five 700 tonne coal feed surge bins or to one of two 75 000 tonne storage piles via travelling stacking conveyors. Drum reclaimers are used to reclaim coal from the storage piles and transfer it to the coal feed surge bins. Drum reclaimers are specified to provide a coal blend of constant size distribution to the feed surge bins.

A 10 000 tonne emergency storage pile is provided for use during plant startup and shutdown or during an emergency until the main coal conveyor from the mine can be shutdown. During normal operation the stockpile is gradually worked off using a front end loader to transfer coal to the conveyor feeding the coal surge bins.

To maintain the coal as dry as possible and to allow for better working conditions for operating and maintenance personnel all outside conveyors are enclosed in heated, lighted galleries and the two 75 000 tonne coal storage piles are enclosed in two 350 m long x 70 m wide buildings. The travelling stacking conveyors, drum reclaimers, feed conveyors and discharge conveyors are inside these buildings. Dust collection and firefighting equipment is provided for the buildings to minimize the risk of fire or explosion.

8.3.1.2 Residue Handling

Solid residue is received from the residue granulation area at a rate of 54.05 tonne per hour by belt conveyor and transferred to a 10 000 tonne residue storage pile via a travelling stacking conveyor or to a solid waste disposal conveyor. A bucket-wheel reclaimer is used to reclaim residue from the storage pile and transfer it at a rate of 22.9 tonne per hour to the Steam Plant.

Coal can be transferred from the main coal storage area to the solid residue storage pile to permit operation of the Steam & Power Generation Plant when the Coal Liquefaction Unit is shut down or operating at reduced rates.

The residue storage pile, travelling stacking conveyor, bucket-wheel reclaimer, feed and discharge conveyors are enclosed in a building.

8.3.2 Area 1200 Steam & Power Generation

The Steam and Power Generation Unit is designed to combust residue from the coal liquefaction process to supply plant steam (in excess of steam produced in the Process Units) and the entire plant power requirements. This unit consists of a steam boiler, three stage turbine generator, ash handling system, and flue gas desulphurization system.

8.3.2.1 Steam Boilers

Solid residue from the Residue Storage area is delivered by belt conveyor to the steam boiler unit. The unit consists of a residue/coal pulverizer, steam boiler, boiler feed water and combustion air preheaters. Alternately, fluidized bed combustion of the solid residue may be considered.

Residue at a rate of 22.9 tonne per hour is pulverized and carried in the primary combustion air stream to the steam boiler. The steam boiler is used to generate 177 600 kg/hr of high pressure (10 400 kPa) superheated steam for the high pressure turbine generator and to superheat high pressure turbine exhaust (1150 kPa) for the intermediate pressure turbine. Effluent streams from the Steam Boiler Unit consist of condensate blowdown, flue gas, and ash. Blowdown from the 10 400 kPa steam drum is transferred to the condensate system for recovery. Flue gas is first cooled in a boiler feed water economizer and combustion air preheaters, then transferred to the flue gas desulphurization unit for SO_2 and fly ash removal. Bottom ash is dumped into ash hoppers and removed by the ash handling system.

8.3.2.2 Power Generation

The steam turbine generator rated at 94.1 MW consists of a high pressure, intermediate pressure and low pressure turbine linked in series to a DC exciter generator to produce 13 800 volt power.

Superheated 10 400 kPa steam is fed to the high pressure turbine and exhausted at 1 150 kPa. The exhaust steam is split into two streams; part returned to the steam boiler to be superheated and the remainder delivered to the plant process steam system. The superheated 1 150 kPa steam is used as feed to the intermediate pressure turbine. Discharge from the intermediate pressure turbine, 450 kPa saturated steam, acts as partial feed to the low pressure turbine with the remaining steam supplied by process steam generators. Part of the 20 kPa exhaust steam from the low pressure turbine is condensed by preheating boiler feed water while the remainder is condensed using air coolers.

Condensate from the boiler feedwater preheaters and the low pressure exhaust steam condensers is mixed with boiler feed water from the condensate recovery system and BFW demineralizers in a deaerator and pumped to the steam boiler.

8.3.2.3 Ash Handling

Bottom ash at 800°C and fly ash at 400°C are gravity dumped from the steam boiler into an ash mix tank. Quench water is sprayed on the ash to cool and slurry it prior to dumping into a submerged chain conveyor.

Ash is dragged from the bottom of the chain conveyor and dewatered to 25% moisture as it proceeds to an ash storage silo. Overflow water from the chain conveyor is drained to settlers where settled sludge is removed and recycled back to the chain conveyor while clarified water overflows to a water surge tank.

Neutralized regeneration water from the BFW demineralizer is used as makeup to the surge tank. This quench water is pumped through air coolers and recycled to the ash mix tanks.

8.3.2.4 Flue Gas Desulphurization

A dry scrubbing system was selected to remove sulphur dioxide and fly ash from steam boiler flue gases. This system has the advantage that neutralized SO_2 and fly ash exit the system as a dry powder, thus eliminating problems associated with sludge disposal. It has the added advantages of eliminating the need for electrostatic precipitators and reheating the flue gas before discharge to the stack.

This system consists of a lime slurrying area, spray drying tower and fabric filter. Two modules are needed to handle the flue gases from this plant, (one operating and one spare). Each module consists of a spray drying tower and fabric filter. A common lime slurrying system is provided.

Lime as CaO is transferred pneumatically from the lime storage area to a lime grinder where it is ground to 74 micron and hydrated to form $Ca(OH)_2$, using neutralized regeneration effluent water from the BFW Demineralizer Plant. The lime slurry is maintained at about a 10% wt concentration and distributed to surge tanks at each module. To prevent settling of solids and plugging of the lines, the slurry is continuously recycled through the delivery lines and returned to the lime slurry tank. The lime slurry is concentrated to 30% wt with recycled salts and fly ash and pumped to the spray towers from the surge tanks with a constant bypass of slurry returning to the surge tanks.

Slurry is dried in the Spray Towers by direct contact with flue gas. A neutralization reaction occurs between the sulphur dioxide in the flue gas and about 80% of the available $Ca(OH)_2$ to form a calcium sulphate salt. Part of the available water reacts with the calcium sulphate to form gypsum ($CaSO_4.2\ H_2O$) as a dry powder and the remainder evaporates.

The flue gases are cooled by spray drying from 160°C to about 15°C above the water dewpoint. The gas stream is reheated with a slip stream of untreated flue gas to 25-30°C above the water dewpoint to prevent water condensation and plugging in the fabric filter. The fabric filter allows intimate contact between gases and unreacted lime and fly ash by forcing the gases through the dry cake, thereby improving overall unit efficiency. Cleaned gases exit the fabric filter at about 65°C and are transferred to the stack using an induced draft fan. The filter cake consisting of gypsum, fly ash and unreacted lime exits the bottom of the fabric filter and is pneumatically conveyed to solid waste storage silos and to solid waste recycle silos. Since the solid waste is alkaline, a portion is recycled to the lime surge tank to increase the operating efficiency of the unit. The remaining solid waste is collected by belt conveyors from the solid waste silos for disposal.

8.3.3 Area 1300 Offsites

The offsites area comprises the storage of raw chemicals and finished product, water pollution control, flare system and solids disposal. The design basis for these areas is outlined in the following sections.

8.3.3.1 Catalyst and Chemical Storage

This section comprises the unloading and storage systems for the coal liquefaction catalyst, lime for flue gas desulphurization, sulphuric acid and caustic soda.

The H-Coal catalyst is assumed to be received in a bulk granular form by rail. Catalyst storage is split into five silos, each silo designed to deliver catalyst to one of the five Coal Liquefaction trains. Each silo has storage capacity for 30 days, totalling 100 tonne. An unloading station is provided for unloading catalyst and transporting the catalyst pneumatically to the storage silos. Catalyst is also conveyed to each Coal Liquefaction train from the silos. Depending on catalyst moisture requirements air or nitrogen may be used as the conveying medium.

Lime for flue gas desulphurization is received by rail car in granular (2 mm dia) anhydrous form (CaO). Two storage silos provide 14 days of storage capacity totalling 350 tonne. An unloading station is provided to unload from rail cars and pneumatically transport the lime to the storage silos. Pneumatic conveying is also used to transport lime to the flue gas desulphurization area.

Concentrated sulphuric acid is used in the BFW Demineralizer Unit to regenerate cation resins on an intermittent basis and in the cooling tower system for pH control. A 30 day supply of 93% sulphuric acid is stored in a 150 tonne storage tank, from which it is pumped to a BFW acid day tank and cooling tower acid day tank.

Concentrated caustic (sodium hydroxide) is used in the BFW demineralizer for the regeneration of anion resins on an intermittent basis, in Light Ends Recovery for the removal of mercaptans from butane and in Phenol Recovery for the removal of phenols from effluent water. A 30 day supply of 50% caustic is stored in a 20 tonne storage tank.

8.3.3.2 Product Tankage and Loadout

Offsites tankage is separated into raw product and finished product storage. Raw product facilities consists of storage for heavy oil for startup purposes and unstabilized naphtha, middle distillate and heavy distillate surge storage. Finished product facilities consist of storage for stabilized naphtha and distillates, propane, butane and ammonia, and syncrude product. Sodium carbonate solution (soda ash) is stored as a finished by-product within the Phenol Recovery Unit battery limits.

An unloading station and 3 000 m^3 storage tank is provided to handle imported heavy oil required to prepare the initial coal slurry for the Coal Liquefaction Unit at startup. The storage tank will provide sufficient heavy oil to operate one Coal Liquefaction train for 18 hours at full operating capacity.

In the event that the hydrotreaters undergo an unscheduled short term shutdown, a surge storage area has been provided to temporarily divert naphtha, middle distillate and heavy distillate flows from the Atmospheric and Vacuum Distillation and Light Ends Recovery areas. A 500 m^3 unstabilized (nonhydrotreated) naphtha storage tank is provided to supply an eight hour surge storage capacity. A 600 m^3 unstabilized middle distillate and a 300 m^3 unstabilized heavy distillate storage tank are also included to provide eight hour storage capacity.

Although the current marketing concept is to sell a blended syncrude product, in reality, naphtha would be segregated from the distillate product for sale separately. Stabilized naphtha, middle distillate and heavy distillate products from the hydrotreaters are stored in 6 000 m^3, 7 200 m^3 and 3 600 m^3 standard API storage tank respectively, providing four days of storage capacity. Products are sampled, tested and approved. Product naphtha is stored in a 36 000 m^3 storage tank providing long term (30 day) storage while product distillate, a blend of middle and heavy distillates, is stored in a 64 800 m^3 storage tank providing 30 day capacity.

Butane and propane from Light Ends Recovery is stored in 4-3 200 m^3 butane spheres and 45-400 m^3 propane bullets respectively providing 30 days storage for each. Anhydrous ammonia by-product from the Ammonia Recovery Unit is stored in 13-400 m^3 horizontal tanks having a 30 day storage capacity. Elemental sulphur is slated and stockpiled in a 2 000 tonne storage pile.

Product loadout facilities are provided for rail shipment of all finished product. Dedicated liquid loadout stations are provided for naphtha, distillate, butane, propane, ammonia and soda ash. Sulphur is loaded into open hopper cars using a travelling belt conveyor and front end loader.

8.3.3.3 Effluent Treatment

Effluent treatment consists of biological treatment of blowdown quench water from the Ammonia Recovery Unit and sanitary treatment of sewage from the raw sanitary sewer.

A total of 69.5 m³/hr of quench water blowdown from the Ammonia Recovery Unit effluent is treated to control quench water solids. Having a BOD loading of about 200 mg/L, the blowdown is mixed with recycle sludge and fed to biological reactors. Phosphoric acid is added as a sludge nutrient and mechanical aerators maintain a uniform mixture and sufficient oxygen during the 24 hour retention time to reduce the effluent water BOD. Overflow from the reactor drains to the clarifier where clarified water is separated and pumped away and sludge is removed and recycled to the biological reactor feed. Clarified water is passed through dual media filters to reduce the total suspended solids prior to transfer to the cooling tower makeup system.

Excess sludge produced by the biological reactor is transferred batch-wise to the aerobic digester. A slow speed aerator provides enough agitation and aeration during a 20 day retention time to destroy a portion of the volatile solids. The resulting sludge is pumped to an effluent centrifuge where it is concentrated to 70% solids. Solids are discharged to a conveyor for solid waste disposal and the filtrate recycled back to the biological reactor.

Sanitary effluent from the raw sanitary sewer enters the biological oxidizer after passing through the comminutor where solids are shredded. Waste water overflows from the biological oxidizer to the sanitary sump, where it is chlorinated and pumped to dual media filters with ultimate discharge to the fire water pond. Sludge from the biological oxidizer is transferred to the sludge digester where it is aerated to reduce the putrescible matter. Sludge from the digester is intermittently pumped to the effluent centrifuge where solids are removed for disposal.

8.3.3.4 Flare System

Seven flarestacks are required for the Coal Liquefaction Complex; one for each Coal Liquefaction train, one flare for the Hydrogen Plant and one general plant flare to serve all other process areas.

The Coal Liquefaction flares are self-supporting stacks rated to accept the H_2-200°C cut off the reactor effluent from one reactor train. The hydrogen flare is a self-supporting stack rated to accept the gasifier effluent stream from one gasifier train. The general plant flare is designed to handle relief and vent vapours from all other process units: Fractionation, Naphtha and Middle Distillate and Heavy Distillate Hydrotreating, Light Ends Recovery, Sulphur Plant, Phenol Recovery, Ammonia Recovery and Product Tankage areas. The flare is a self-supporting stack designed to handle the H_2-200°C cut off the fractionator overhead vapour as the maximum flow rate.

Vents from the Hydrogen Plant are flared directly while process vents from the Coal Liquefaction and other process units pass through knockout drums. In addition, blowdown separators are installed upstream of the Coal Liquefaction knockout drum to knock out any slurry or solids carried over in the vent vapours.

8.3.3.5 Solid Waste Disposal

Solid waste must be removed from the ash handling and flue gas desulphurization units continuously and from the effluent treatment unit intermittently. A total of 62 000 kg/hr of solid waste is collected and transported by belt conveyor to an acceptable disposal site. Excess solid residue from the liquefaction reaction must be disposed of in an environmentally acceptable manner -- incineration is considered an acceptable option.

8.3.4 Area 1400 Utilities

The Utilities Area comprises the air, steam, condensate, BFW, cooling water, potable water supply and fuel gas system. A brief description of the design basis for these areas is outlined in the following sections.

8.3.4.1 Air Systems

Compressed plant and instrument air is provided by two oil free compressors with a common spare. All three compressors are capable of supplying air at 800 kPa. In addition, an air dryer is provided on the instrument air system to reduce the dewpoint to -40°C.

8.3.4.2 Steam & Condensate Distribution

Four steam pressure levels were selected for plant use: 10 400 kPa, 4 250 kPa, 1 150 kPa and 450 kPa. High high (10 400 kPa) pressure steam produced in the Steam Plant, Coal Liquefaction waste heat boilers and Residue Gasification waste heat boilers is used exclusively to drive the high pressure power generation turbine. An intermediate stream is withdrawn from the high pressure turbine and blended with process steam produced in the hydrotreaters and oxygen plant to feed the high pressure (4 250 kPa) header. High pressure steam consumers include the coal liquefaction and light ends recovery area.

Excess exhaust steam from the high pressure (10 400 kPa) power turbine is added to process steam from the light ends recovery, residue gasification and oxygen plant areas to feed the intermediate pressure (1 150 kPa) steam header. Intermediate pressure steam users include coal liquefaction, fractionation, hydrotreating, light ends recovery, sulphur plant and ammonia recovery.

Low pressure (450 kPa) steam for plant use is generated in coal liquefaction and oxygen plant waste heat boilers. This steam is fed to the residue gasifiers and used throughout the complex for steam tracing and building heating.

The condensate collection and blowdown system consists of a series of blowdown and flash drums to produce atmospheric condensate and blowdown. The blowdown is used as cooling tower makeup while the condensate is deaerated and distributed.

8.3.4.3 Raw Water Treatment

Raw water is supplied to the plant from remote pumping stations drawing from local wells and the Elk River. Raw water is first passed through a sand type gravity filter to remove residual solids and distributed to the cooling tower service water and quench water systems for use as filtered water and to the BFW plant for demineralization.

8.3.4.4 Potable Water

The Potable Water Treatment Unit is a complete package unit designed to purify and disinfect raw water. Potable water is stored in a 15 m^3 storage tank to provide a 10 hour supply to the complex at design flow rates.

8.3.4.5 Boiler Feed Water

The BFW Demineralizer Plant receives 178 m^3/hr of filtered water from the filtered water supply pumps to provide demineralized water to the BFW system. The BFW Plant consists of two parallel modules, each producing 50% of the total BFW requirements. Each module consists of a pretreatment carbon filter to remove trace organic compounds in the water, a primary weak cation column, primary strong cation column, degasser, primary weak anion column, secondary strong cation column and primary strong anion column. The moderate hardness of the raw water allows the use of demineralizer resins without the need for a softening system upstream of the demineralizers.

Of the 178 m^3/hr of boiler feed water produced, 150 m^3/ hr is stored in a 10 000 m^3 storage tanks providing a three day supply of BFW and 28 m^3/hr used as regeneration dilution water. Concentrated 93% sulphuric acid is diluted with BFW and used to regenerate the cation beds and 50% caustic soda diluted with BFW to regenerate the anion beds. The neutralized effluent water is pumped to ash handling as makeup water, to flue gas desulphurization to slurry lime for sulphur dioxide removal, and to the effluent holding pond as effluent water.

8.3.4.6 Cooling Tower System

Due to the shortage of water in the plant location, air cooled heat exchange has been maximized to reduce the overall complex cooling water demand. With this policy the cooling water demand for the entire complex is 13 600 m^3/hr.

Cooling water is supplied to the Coal Liquefaction, Atmospheric and Vacuum Distillation, Hydrotreating, Light Ends Recovery, Sulphur Plant, Hydrogen Plant, Oxygen Plant and Ammonia Recovery Units with blowdown delivered to the fire water pond as effluent. The cooling tower cools 13 600 m^3/hr of water from a design inlet temperature of 38°C to an outlet temperature of 27°C. Estimated water losses due to evaporation and drift are 540 m^3/hr.

Makeup cooling tower water is composed of condensate blowdown from the condensate recovery system, effluent treatment water and filtered water from the raw water treatment system. The cooling water quality is maintained by the addition of sulphuric acid from the cooling water acid pump for pH control, addition of inhibitors to prevent scale deposition and corrosion of cooling tower piping, and biocide injection to prevent algae and slime formation.

8.3.4.7 Effluent and Fire Water System

Blowdown from the cooling tower and sanitary water discharges into the fire water pond then overflows to the effluent water holding pond. Neutralized effluent from the BFW Plant bypasses the fire water pond and is discharged directly to the effluent holding pond as this water is of poor quality. The fire water pond is separated from the effluent water pond, thus allowing the effluent holding pond to be drained without affecting the fire water supply. The effluent holding pond is sized to hold 35 000 m^3 of effluent water to allow a two week holdup.

Fire water is provided to a plant fire water main loop and a fire water high pressure loop by fire water pumps at the fire water pond and the raw water pipeline. Jockey pumps are used to maintain the system pressure. The fire water pond having a capacity of 3 600 m^3 can supply water to the fire loop for five hours without the raw water fire pumps in operation. Gasoline storage tanks are provided to fuel spare fire water pumps.

Socio-Economic Considerations

9.0 SOCIO-ECONOMIC CONSIDERATIONS

9.1 REGIONAL PROFILE OF THE EAST KOOTENAY REGION

9.1.1 Introduction

The Regional District of East Kootenay (RDEK) covers the south-east corner of the Province of British Columbia and is enclosed by Alberta on the east, and the U.S.A. on the south. It's northern boundary lies between Radium Hot Springs and Golden, B.C. The western boundary is formed by the Purcell Mountain range. The region is comprised of three separate valleys, each having its own economic stimulus with Cranbrook as the major regional service center. The project area lies within the Elk Valley sub-region which contains the city of Fernie, the District of Sparwood and the Village of Elkford.

The climate within the region varies widely between different locations, however, it generally can be characterised as continental and semi-arid in nature, with extreme temperature ranges and low precipitation in the main valleys.

9.1.2 Demography

In 1979, the population within the Region was estimated to be about 50 000, an increase of about 10 000 over the 1971 figure and an increase of 3 500 over the 1976 total of 46 450.

In 1976, 48.4% of East Kootenay's population was younger than 25 years of age compared to 42.8% for British Columbia. The school age population was 14 075 or 30.4% which was higher than the Provincial figure of 26.7%. The in-migration of workers has lead to a slight male predominance in 1976, 51.7% of the regional population being male compared with 49.9% for the Province as a whole.

In the past 10-15 years, population growth in the RDEK has been somewhat fueled by natural growth, but more importantly, by in-migration associated with the development of Elk Valley coal resources. The population distribution among the major settlements in 1976 and 1979 is shown in Table 9.1. In 1979, Cranbrook had over one-third and the three Elk Valley communities had more than one-quarter of the regional population.

It is anticipated, as illustrated in Table 9.2, that by 1990 the total population of the RDEK will be approximately 110 000. The combination of Elkford, Sparwood and Fernie will account for 37 500 or about 35% of this total. By 1990 the community of Elkford is projected to grow to 11 000 from its 1981 population of 3 400, which represents an increase of approximately 225%. Much of this growth will be induced by the manpower requirements estimated by the various coal mining companies operating in the RDEK over the next several years along with related growth in the service and commercial sections of the economy.

According to a recent analysis carried out on behalf of the Economic Development Commission of the RDEK, entitled Economic Overview for 1982, rapid growth will be experienced in the region up to 1985 during which time many of the proposed mines will be brought into production. This rapid growth is predicted to extend up to 1990 because a number of additional mining and related major projects are planned for development after 1985.

9.1.3 Labour Force

Past trends in labour force availability are illustrated below:

1971 - 16 510
1976 - 19 925
1979 - 25 000 (estimate only)

In 1977, the largest employment sector of the economy was Community and Personal Services employing 20.9% of the community labour force which included tourism, health, education and a variety of professional services. The mining industry employed 19.1% while 13.5% were in the retail and wholesale trade. Manufacturing, primarily forest processing, employed 11.0% and transportation communications and utilities employed 9.5% of the labour force. (See Table 9.3).

Traditionally, unemployment rates have shown large fluctuations, being significantly below the provincial average in buoyant periods and well above the latter when the economy experienced downturns.

During the winter of 1975/76, the Cranbrook CMC area recorded 7.8% unemployment. Male unemployment is concentrated in construction, transportation and processing. Female unemployment is largest in the clerical, sales and service occupations. In the Cranbrook CMC area, female unemployment insurance claimants formed 50% of total claims, although female represent only 30% of the labour force (1976 data).

The Fernie Canada Manpower office advised that there is a growing shortage of heavy duty mechanics, millwrights and industrial electricians. The shortage of equipment operators experienced in earlier years appears to have been eliminated.

Personal incomes are equal to or slightly above the provincial average, but household incomes lie slightly below those of the province as a whole.

9.1.4 Economy

Mining Industry

Employment and production in the region's mining industry has been growing since the early 1970's.

Mechanization in the mines is expected to reduce employment levels, but development of new mines will increase employment opportunities.

There are small deposits of metallic minerals throughout the Region, which have been mined at various times. These include gold, silver, copper and platinum. Non-metallic mining of commercial significance include sand and gravel, gypsum and barite.

Coal is currently the most important of the region's mineral resources. The coal deposits are large enough to be of national significance and are found along the western slopes of the Elk Range in the Rocky Mountains, throughout the Crowsnest Pass area, all along the eastern side of the Lower Elk Valley and less extensively in the Flathead Valley. Metallurgical coal is mined at three open-pit sites. Two underground mines just east of Sparwood are also in operation. At present, five more mines are in the planning and development stages.

Forestry

There are two Timber Supply Areas in the region at Invermere and Cranbrook. The total Allowable Annual Cut for the provincially-managed forests is approximately 77% of the forest area and is 1 822 000 m³. Spruce is the predominant species and constitutes the highest percentage of the total timber supplied (37.6%).

Agriculture, Parks and Park Reserves

The region's topography and climate do not favour agricultural activity. A 1976 Census of Agriculture revealed that only 2.6% of the total land area (73 750 hectare) is classified as farm land. About 80% of this total is used for grazing and fodder crops with the remainder being under cultivation.

The region's spectacular scenery and attractive environment are resources for public enjoyment in several major parks. The biggest attraction within the region is the Kootenay National Park, with its Radium Hot Springs. This park is 140 750 hectares with 116 000 hectares of other provincial parks.

Tourism and Recreation

The role of the tourist industry has been steadily growing over the last decade. The spectacular scenery and invigorating climate have served to popularize outdoor recreation activities throughout the year for both local residents and visitors. The number of visitors in 1978 was 820 000, a growth of 8-10%. The revenue from tourism in the region for 1978 was estimated to be $170 million.

Other Industries

Other industries in the region include manufacturing, commercial services, transportation and utilities.

Manufacturing industries are dominated by forest processing, sawmilling and pulping. Additionally, there is a small amount of mineral and secondary forest processing. The commercial and service industries are concentrated in Cranbrook and serve a regional market of 50 000 people. Fernie is the major service centre for the Elk Valley market of 15 000 people. The largest regional employers in the transportation and utility field include C.P. Rail, with 350 employees, B.C. Hydro with 150 employees, B.C. Highways with an average of 200 employees and B.C. Telephone with 250 (approximately) employees within the region.

9.1.5 Infrastructure and Community Facilities

The RDEK is served by three (3) major highways, (provincial numbers 3, 93, 95/95A). The southern transmountain route along Highway 3 is the most heavily used. All sizeable communities in the region are located on these highways with the exception of Elkford which is 25 km north of Highway 3.

The Canadian Pacific Railroad is routed across the region and is mainly used commercially for the transportation of grain and coal.

The major airport in the region is the Cranbrook-Kimberley airport administered by the City of Cranbrook and served by Pacific Western Airlines. Commercial passenger and freight levels in and out of the airport in 1978 were estimated at 96 000 passengers and 1.0 million pounds of freight. A small local airport is situated near Elko.

Scheduled bus services are run by Greyhound lines of Canada and Cranbrook-Golden Bus Lines.

The utilities in the region are administered by the following authorities: electricity by B.C. Hydro and Power, natural gas by Columbia Natural Gas, and telephone by B.C. Telephone. One of the largest consumers of electricity is the coal mining industry.

The region's medical facilities are administered by the Regional District of East Kootenay Hospital Board. There are several hospitals, public health clinics and nursing homes in the region.

The primary and secondary education is divided into four school board districts. The Fernie district has ten elementary and four secondary schools, which presently are full to capacity. Post secondary education is under the jurisdiction of the East Kootenay Community College, offering academic, vocational and continuing education in a wide variety of fields. A new $7.5 million college campus is presently under construction in Cranbrook.

Housing in the area is in short supply for rental purposes, but the level of company subsidies to the employees maintains rentals at relatively low cost. Housing and industrial land is available in the region and construction costs in 1979 dollars were $40/sq ft for frame buildings and $60/sq ft for institutional and commercial buildings.

9.1.6 Regional Economic Potential

Current prospects for rapid expansion in all sectors over the next decade appear to be more than favourable. These prospects will depend on the growth of the industrial and resource bases. Development for the immediate future (1-5 years) will be concentrated in the Elk Valley and Columbia Valley sub-regions, stimulated by coal mining and tourism, respectively. The expansion of these basic economic sectors and creation of new jobs will attract population to the communities of Fernie, Sparwood and Elkford. A share of the spin-off jobs in the service sector is expected to be concentrated in the central area, especially Cranbrook.

This growth has already been anticipated, but many local infrastructure services must be expanded if this growth is to be accommodated without causing deterioration in local servicing standards.

Coal reserves will permit the East Kootenay Regional District to grow provided existing coal markets are expanded and new markets created. Many new projects are proposed and also expansion of existing projects are underway at the Fording and B.C. Coal facilities. B.C. Hydro is currently carrying out a study to generate electricity from coal waste. Total investment in the coal industry related infrastructure over the next decade, if all these projects proceed, would be at least $2.0 billion, creating 4 500 jobs which is 2.5 times the present level of employment. This demand would outstrip the regional market and training facilities to provide qualified labour.

The forestry industry within the region is unlikely to grow due to the decline in resources over the past two decades. The agricultural industry is unlikely to grow, due to the unfavourable conditions already outlined in the region.

There is substantial commercial, residential and industrial development planned and underway particularly in Fernie, Sparwood and Elkford.

TABLE 9.1

POPULATION DISTRIBUTION 1976-1979

	1976	%	1979	%
Regional District of East Kootenay	46 450	100.0	50 099	100.0
Columbia Valley:	5 421	11.6	N/A*	
Invermere	1 194	2.5	1 800	3.5
Central Area:	28 308	60.9	N/A	
Cranbrook	13 510	29.0	16 500	32.9
Kimberley	7 111	15.3	7 100	14.1
Elk Valley:	12 415	26.7	N/A	
Fernie	4 608	9.9	6 000	11.9
Sparwood	4 050	8.7	4 300	8.5
Elkwood	1 873	4.0	3 000	5.9

* Not available

Sources: Statistics Canada
B.C. Research
Remaining 1979 populations are based on Municipal estimates.

TABLE 9.2

REGIONAL DISTRICT OF EAST KOOTENAY

POPULATION PROJECTIONS

These projections are based on estimates given by the coal companies as to their manpower requirements over the next several years. We have attempted also to provide for the rapid growth expected in the commercial and service sectors.

	1981	1985	1987	1990
Elkford	3 400	7 500	8 792	11 000
Sparwood	6 000	8 000	10 000	12 000
Fernie	7 500	10 000	12 000	14 500
Cranbrook	17 100	23 000	28 000	32 000
Kimberley	8 000	9 000	10 000	10 500
Invermere	3 000	6 000	7 500	9 000
Rural: including Radium, Canal Flats, Wasa	15 000	17 000	19 000	21 000
	60 000	80 500	95 292	110 000

The fastest growth is anticipated up to 1985, when the new mines will all be in production. Also during this period, we will see the fastest growth in the commercial and service sectors. Mines and other very large projects planned for development after 1985 are cause for the balance of escalation to 1990.

Source: Economic Overview for 1982 - Economic Development Commission Regional District of East Kootenay.

TABLE 9.3

LABOUR FORCE ESTIMATES BY INDUSTRY AND SUB-REGION

	Elk Valley*		Central Area		Columbia Valley		Total Regional District	
	June 1 1979	(%)	1977	(%)	1976	(%)	1977	(%)
Agriculture	50	(0.8)	250	(2.0)	80	(3.2)	375	(1.7)
Forestry	150	(2.2)	225	(1.8)	220	(8.7)	550	(2.6)
Mining	3050	(45.5)	1125	(8.9)	100	(4.0)	4100	(19.1)
Manufacturing	625	(9.3)	1350	(10.7)	550	(21.8)	2375	(11.0)
Construction	400	(6.0)	1000	(8.0)	130	(5.1)	1650	(7.7)
Transportation, Communication & Utilities	350	(5.2)	1575	(12.5)	110	(4.3)	2050	(9.5)
Trade	500	(7.5)	2325	(18.5)	230	(9.1)	2900	(13.5)
Finance, Insurance and Real Estate	100	(1.5)	475	(3.8)	65	(2.6)	650	(3.0)
Community, Business & Personal Services	1000	(14.9)	2775	(22.1)	800	(31.7)	4500	(20.9)
Public Administration	150	(2.2)	575	(4.6)	90	(3.6)	850	(4.0)
Unclassified ***	325	(4.9)	900	(7.2)	150	(5.9)	1500	(7.0)
Totals	6700	(100.0%)	12575	(100.0%)	2525	(100.0%)	21500	(100.0%)

* Elk Valley figures for School District No. 1, not including Jaffray.

*** Includes first job seekers.

Note: These estimates are generated by CEIC on an "as required basis" and hence no consistent data for a single year are available.

Source: Data provided by Economic Services Branch, Canadian Employment and Immigration Commission, Kamloops.

9.2 THE PROJECT FROM A SOCIO-ECONOMIC PERSPECTIVE

Construction and operation of the coal liquefaction facility will provide a major economic stimulus to British Columbia, the RDEK and the community of Elkford.

At the provincial level, the project will utilize existing labour resources that at present appear to be idle. It will also generate demands for those skills which appear in short supply. This will raise a number of implications for the training policy related to skilled trades and its ability to satisfy demands in the long term. In addition, the project will stimulate the manufacturing and construction material sectors of the provincial and regional economy due to demands for equipment and materials. In those areas where existing resources can satisfy the demands of the project, it is unlikely that any socio-economic disruption will result. Alternatively, where existing resources are in short supply and project demands cannot be readily satisified, then socio-economic disruptions may result.

In order to retain the maximum number of benefits within the region, the RDEK will have to improve the capabilities of the service and supply sectors of its economy (Elco Mining 1978). Population growth within the District occurs predominately in association with coal mining developments at the eastern fringe of Cranbrook's service area in the Elk Valley Sub-region.

This area is situated beyond the service limits of the District's transportation and settlement framework, which centres around the community of Cranbrook. Consequently, the project raises the need for creating a sub-regional service centre which can become the focus and provide support for accelerated population growth to be experienced in the area.

At the local level, it can be anticipated that the community of Elkford will be severely strained in terms of its public, commercial, services and supply capabilities. Consequently, realization of the project will require a significant expansion of the community's infrastructure including community services. However, such a change will be highly beneficial to the community as well as to the region. It will permit Elkford to move beyond its present level of development towards a more diversified economy with the hope of improving stability and resident satisfaction. It will provide a centre for the Elk Valley Sub-region permitting greater efficiency in the delivery of social services as well as maintaining population stability (Suzanne Veit, 1979).

Currently within the region major employment opportunities exist in the primary sector of the economy, particularly in mining and forestry and to a lesser extent, in tourism. Development of the liquefaction facility will provide a new alternative employment opportunity within the region. Furthermore, the facility will stimulate the development of a diverse range of industries related to servicing and maintaining the facility. Such development will permit the region to diversify its economy and provide a greater variety of employment opportunities for its residents.

Finally, the longer term effect on recreational resources from the expanded population in the Elk Valley Sub-region is a major socioeconomic aspect of the implementation of the project (Elco Mining, 1978). The Kootenay and Columbia Valleys are said to be already straining under the demands placed upon them by the growing populations of the East and West Kootenay Districts, the increasing use by Albertans and the growing number of international tourists.

9.3 MANPOWER REQUIREMENTS AND AVAILABILITY

Construction of the liquefaction facility is estimated to take 3 1/2 years for completion during which time labour will peak at approximately 5 000 workers. It is anticipated that the construction phase will begin in mid-1985 with start-up scheduled for the first quarter of 1989 and full operation by 1992. The operation of the facility will require approximately 1 000 workers. At this time, precise requirements for labour during the construction and operation phases are still evolving with decisions regarding process selection and the subsequent refinement of engineering, operational procedures and equipment selection.

Since construction of the facility is estimated to begin in 1985, it is difficult to predict with certainty the labour availability at that time. However, availability will be influenced by several factors including:

i) competing demands within the Province
ii) competing demands in Western Canada
iii) size and distribution of labour skills

The range of developments underway and proposed within British Columbia as of October 1981 is summarized in Table 9.4. These and other unidentified projects are expected to create a continuing strong demand for construction workers in the province.

It is considered unlikely that all of the projects will begin at the same time. In fact, it may be possible to refine the construction schedule in such a manner that the liquefaction project will fit into a labour window created by the variable peaking of other projects. This will require greater detail regarding skills and numbers of workers as well as more precise information regarding the construction scheduling of the other projects.

At present, much uncertainty exists regarding the status of some of the mega-projects in Western Canada. A more detailed appreciation of the status of these projects will be necessary in order to predict their influence on labour availability in Western Canada.

Previously it was estimated by Alberta's Advanced Education and Manpower Planning Secretariat (1976) that the demand for construction workers was expected to peak in 1976-77 and to display a relative decline into 1978-79. Nevertheless, should several of those projects be given the go-ahead by 1985, then they could provide significant competition for labour in Western Canada.

In 1978 Elco Mining submitted a Stage II Report discussing the economic benefit and socio-economic effects of developing the Elk River Coal Project located approximately 30 km from Elkford. They indicated that approximately 1 100 construction workers would be required around 1980 to develop a mine with assumed start-up in 1982. At that time, available evidence indicated that no major difficulties would be experienced in obtaining the required work force.

In 1980 Kaiser Resources issued a Stage I Socio-Economic Assessment report, indicating that approximately 600 permanent jobs would be created to operate the proposed Greenhills mine. Concern was expressed regarding the availability of qualified labour.

Recently, the Province has completed a skill shortage survey for the construction industry, the results of which have not yet been released. However, one of the tentative conclusions suggests that there is likely to be some shortage of iron workers, piping trades, sheet metal trades and elevator mechanics in the period until 1984.

Manpower supply for the operational phase will likely emphasize sources within the Western Provinces. While not seeking to limit the free movement of labour within Canada, local workers are likely to be more stable and more familiar with working and cultural conditions in the West. Localization reduces recruiting costs and raises the relative efficiency of recruiting programs. Emphasis on local hiring permits a better appreciation of and adjustment to the local manpower supply in terms of training programs.

In general terms it is believed that the supply sources of manpower will include:

i) petrochemical industries in B.C. or Alberta
ii) heavy industry in Western Canada
iii) other industry in Western Canada
iv) new entries into the work force including institutional graduates and participants in Government and special programs
v) women and unemployed welfare recipients involved in on-the-job training programs and other local educational services

A recent critical trades study carried out by the Ministry of Labour for British Columbia indicated that a shortfall of approximately 5 000 workers would occur by 1984 (see Table 9.5). In order to overcome this shortfall, it was estimated that a 150% increase in the present number of apprentices would be necessary. (Ministry of Labour, 1981).

The top ten trades included millwrights, heavy duty mechanics, industrial electricians, machinists, carpenters, steam and pipefitters, steel fabricators, painters, auto mechanics and ship and boat builders. These trades accounted for 80% of the labour stock and almost 90% of the shortages. The top four trades mentioned above represent 60% of the labour stock and 55% of the shortfall.

The traditional construction trades (carpenters, sheet metal workers, steel fabricators, etc.) are found in large numbers in the industrial and public administration sectors. Although these trades represent a relatively large proportion of forecast shortages they cannot be considered as a major imbalance until the situation within the construction industry is identified.

Forecasted shortfalls of machinists represent a 50% increase in present stocks. Reduction of the shortfall would require an estimated 400% increase in the present number of apprentices.

While large numbers of apprentices are being trained in the heavy duty mechanic and industrial electrician trades, it appears that a doubling would be necessary to prevent a shortfall. Steel fabricators also indicate large shortfalls of labour for 1984. However, there is some indication that employers are finding welders adequate for the job; consequently, these shortfalls may be overestimated.

Since operation of the facility is estimated to begin in 1989, with full operation by 1992, it is difficult to predict with any certainty labour availability at the time. However, based on population projections of 110 000 for the region by 1990, it is anticipated that a total labour force of approximately 50 000 will exist. As only about 1 000 workers will be required for this project, no serious strain on the regional labour force is anticipated.

TABLE 9.4

LARGE ENGINEERING PROJECTS ANTICIPATED IN B.C. IN THE 1980's (1)

Company Name	Type of Project	Location	Cost (Millions)	Manpower Construction (Peak)	Manpower Operations	Timing	Status
COALMINES							
Line Creek	Coalmine OP	Crows Nest	150-200	300	440	1980/82	Underway
B.C. Coal Greenhills	Coalmine OP	Sparwood	280	400	660	1981/83	Underway
B.C. Coal - Sparwood	Coalmine OP/UG Exp.	Sparwood	-	-	-		Pre-Feasibility
B.C. Coal - Hosner Wheeler	Coalmine UG	Sparwood	200	-	600-800		Dormant
Rio Algam - Sage Creek	Coalmine OP	S.E. - B.C.		380	510	1982/84	Under Review
ELCO - Elk River	Coalmine OP	Elk River Valley	150 (?)	890	1 320	1981/85	Under Review
Denison - Quintette	Coalmine OP	N.E. Coal Block	700	1 800	640 (-1 460)	1981/83	Underway
Fording - Eagle Mountain	Coalmine OP/Ext.	Fording River	115	165,000 Manhours	150	1981/83	Under Review
Teck - Bullmoose	Coalmine OP	N.E. Coal Block	220	462	500	1981/83	Underway
B.P. Sukunka	Coalmine UG	N.E. Coal Block	300	1 100	430 (-1 160)	1981/82	Stage 2 Approved
Petrocan - Monkman	Coalmine OP	N.E.	216	1 600 (total manyears)	861	1982/85	Under Review
Teck - Burnt River	Coalmine OP	N.E.	n.a.	150	265	1982+	Under Review
Utah - Carbon Creek	Coalmine OP/UG	N.E.	n.a.	636			Under Review
Semper - Willow Creek	Coalmine UG	N.E.	n.a.		61	1983+	Under Review
Quinsam	Coalmine OP	Campbell River	70-80	177	200	1983	Under Review
Norco - Bowron River	Coalmine UG	Central B.C.	78	250	200	1982/85	Under Review
INDUSTRIAL							
Canadian Cellulose	Pulp Mill Expansion	Castlegar	360	800		1982/85	Considered
Cominco	Modernization	Trail	350	500		1977/84	Underway
Western Forest Products	Pulp Mill Expansion	Woodfibre	205	400		1981/84	Underway
Port of Prince Rupert	Grain Elevator Coal Terminal	Prince Rupert	140 95	150 300		1982/84 1981/83	

TABLE 9.4

LARGE ENGINEERING PROJECTS ANTICIPATED IN B.C. IN THE 1980's (1)

Company Name	Type of Project	Location	Cost (Millions)	Manpower Construction (Peak)	Manpower Operations	Timing	Status
INDUSTRIAL (cont'd...)							
C.P. Rail	Tunnel	Rogers Pass	500	500-600		1982/86	Planned
Port of Vancouver	Roberts Bank Vancouver Expansion	3 Terminals	50 100				Proposed Proposed
BCR	Spurline	N.E. Coal Block	315				
CNR	Upgrading Pr. George - Pr. Rupert		238				
METAL MINES							
Cominco - Valley Copper	Copper/Moly. Mine OP	Highland Valley	750	1 400	846	1982/83	Approved
Noranda - Goldstream	Copper/Zinc/Silver Mine	Revelstoke	62	250	185	1981/82	Approved/ Underway
Consolidated Cinola	Goldmine OP	Queen Charlotte Islands	170	300	267	1982/83	Under Review
Cominco	Exp. & modernization	Trail	425			1977/84	Underway
POWER GENERATION							
B.C. Hydro - Hat Creek	Thermal Coalmine OP	90 km W. of Kamloops	5 000	3 200	660	1983/91	Under Review
B.C. Hydro - Site C	Hydro Electric Project and Transmission	Peace River	1 086.2	2 045		1982/88	
B.C. Hydro - Meagher Creek	Thermal			5 000			
B.C. Hydro - Liaird	Hydro Electric Project		10 900	3 500			
B.C. Hydro - Murphy Creek	Hydro Electric Project		954	1 340		1984/89	Under Review
B.C. Hydro - Revelstoke	Hydro Electric Project		1 200	2 435		1977/84	Underway
B.C. Hydro (Alcan) - Kemano II	Hydro Electric Project		2 500	750			Under Review
B.C. Hydro - Stikine/Iskut	Hydro Electric Project		5 506	3 500			
OTHER ENERGY PROJECTS AND RELATED							
Foothills B.C. - North	Gas Pipeline	N.E. B.C.	100-1 500	6 000		1982/85	Under Review
B.C. Hydro or Westcoast	Gas Pipeline	S. Mainland to Vancouver Island	125	400		1983	Renewed Review

(1) Preliminary revised draft (October 1981)

SOURCE: Ministry of Industry and Small Business Development - 1981.

TABLE 9.5

TRADES ACROSS ALL INDUSTRIES SURVEYED
(Trade Summary)

Trade	Estimated Present Stocks 1980	Estimated Present No. of Apprentices	Estimated Shortfall by 1984	Estimated % Increase In Stocks Required
Millwrights	4300	770	750	20%
Heavy Duty Mechanics	2800	662	700	25%
Indus. Electricians	2200	431	450	20%
Machinists	1600	206	800	50%
Carpenters	1600	165	550	35%
Steam & Pipefitters	1200	249	150	15%
Steel Fabricators	900	124	550	60%
Painters	650	26	300	45%
Auto Mechanics	600	42	175	30%
Ship & Boatbuilders	450	27	n/a	--
Sheet Metal Workers	450	80	200	45%
Joiners	350	20	150	45%
Indus. Instru. Mech.	350	107	50	15%
Electronic Tech.	325	43	n/a	--
Plumbers	150	4	70	50%
Moulders	100	14	10	10%
Cement Masons	75	3	10	15%
Refrigeration Mech.	75	8	20	30%
Electrical Const.	60	12	10	20%
Ironworkers	60	51	25	40%
Diesel Mechanics	50	8	n/a	--
Patternmakers	50	8	10	20%
Boilermakers	40	9	15	40%
Bricklayers	40	9	5	15%
Auto Body Repairmen	30	1	5	20%
Roofers	30	4	5	20%
Sprinkler Fitters	5	5	--	--
Glaziers	5	0	--	--
TOTAL	18 545	3 088	5 010*	

NOTE: Of the firms responding to the survey, no tradesmen were found in the drywalling, boilermaker (erection), floor covering or wall and ceiling installation trades.

*Totals differ from industry summary due to rounding.

SOURCE: Government of British Columbia, Ministry of Labour.

9.4 SOCIO-ECONOMIC IMPLICATIONS OF THE CONSTRUCTION AND OPERATION OF THE LIQUEFACTION FACILITY

During the construction phase it is anticipated that all of the labour requirements will be satisfied within Western Canada, the majority of which will be satisfied by resources in British Columbia and to a lesser extent, Alberta. Temporary accommodations for the construction workers will likely be in the form of a construction camp since local resources are quite limited in this regard. The total direct and indirect employment estimated to result from the construction of the facility will be approximately 13 000 within British Columbia. (Statistics Canada, 1974).

During the operation phase, the coal liquefaction facility will generate an estimated 4 000 direct and indirect employment opportunities within British Columbia. (Statistics Canada, 1974). At the local level, assuming a similar multiplier to the one used by Elco Mining in their submission, the Village of Elkford will experience an estimated population increase of 3 500. This includes workers directly employed in the facility, their families and those required to provide additional services to the area.

In 1981, the Village of Elkford had a population of 3 400. Currently, Elkford is realizing a strong growth period. The population of the community increased substantially between the years 1971 to 1981. Upon incorporation in 1971 the population was a mere 220. Elkford currently projects, after recent announcements of potential developments for the area, that the 1985 population will be 7 500, the 1987 population 8 800 and the 1990 population, 11 000.

The influx of approximately 3 500 people associated with the development of the coal liquefaction facility into Elkford around 1990 will effect the level of services provided by the village. Elkford, realizing the demands that increased population places on services, is currently expanding its service base to accommodate the projected increases. However, current expansion plans may not be adequate as projected population growth between 1987 and 1990 is only 2 200.

Basic services and needs required by the additional population due to this coal liquefaction facility include:

1) approximately 1 300 additional residential housing units

2) additional teachers for the expected new students

3) increased municipal and physical services to handle the additional residential housing units

4) adequate recreation, hospital/clinic facilities and associated public health services

5) increased federal and provincial government services

6) expansion of commercial retail shopping, recreation and entertainment facilities

7) expansion and upgrading of local and regional road networks and railway service

Many of these services will already exist in Elkford by the 1990's, therefore, the influx of 3 500 additional people may only require the expansion of existing services. Requirements such as accommodation, education, health and municipal services will require advance planning in order that the services are available upon the arrival of the additional people.

Environmental Considerations

10.0 ENVIRONMENTAL CONSIDERATIONS

10.1 OVERVIEW

For the present study, it is assumed that the coal liquefaction complex is located in the B.C. Coal Property area at the Greenhills Mine site. The development of such a large-scale project may have effects on the natural and physical components of the environment within the project area. However, many of the potential impacts can be avoided, eliminated, or minimized if they are identified prior to the construction phase and mitigation measures are incorporated into the construction method and engineering design.

The likely impacts of the project on the environment can be categorized into three types: short-term impacts primarily occuring during the construction phase, long-term impacts usually associated with the operation phase, and residual impacts indefinitely remaining after the plant decommissioning.

This section discusses the potential environmental impacts of the project according to the above three categories; it also describes the control measures designed on the basis of best available control technology to eliminate or minimize these impacts.

The analyses of potential environmental impacts are carried out with reference to the present process plant configuration, the "Greenhills Coal Project Stage I Environmental Assessment" report, and the environmental baseline data given in the report "Elk River Coal Project Stage II Report, Volume III - Environment". The Elk River Coal Project at Elk River Valley, British Columbia, is approximately twenty (20) kilometres north of the Greenhills Mine site.

10.2 POTENTIAL IMPACTS

10.2.1 General

While the site specific environmental data are not fully available at present, based on a general appreciation of the project area, it appears that the project does not generate any unresolvable impacts on the environment, of which the land is already commited to industrial use.

The various components of potential environmental impacts are discussed briefly in the following sub-sections.

10.2.2 Topography and Land Use

Construction

Preparation of the site for the coal liquefaction project involves the removal and replacement of a large quantity of earth materials. Subsequently, structured buildings, storage ponds, access roads and other facilities are constructed. All of these activities would result in unavoidable changes in land topography.

During construction, the land adjacent to the project site would also be disturbed by the equipment/vehicle movement and other related activities. This disturbance is considered relatively minor and of short-term.

The land use of the project site, however, would not be affected by the construction of the complex, since presently the Greenhills Mine site is designated for industrial development use.

Operation

During the operation phase, there would be increasing wear on the surrounding landscape, caused by the activities of men and machinery. As time progresses, more acreage of land would be taken up for solid waste disposal/landfill, storage piles and access roads.

Post-Operation

Once operation ceases, the impacts on the topography of the site would be minimal. The area directly occupied by the coal liquefaction complex may or may not return to other uses, depending upon the type of reclamation planned at the outset of the development of the project.

10.2.3 Soils

Construction

The construction of large-scale complex may have an adverse impact on the soils in the project area: erosion of the soils around the site.

Construction activities at the project site involve the removal of vegetation and the disturbance of soils. If proper precautions are not taken, the disturbed soils would be exposed to wind, precipitation and runoff; slopes would be altered, drainage courses blocked, excavated soils stockpiled without covers, and the physical properties of soils changed. The result would be an extremely high rate of soil erosion and the loading of nearby streams with large volumes of sediment. By incorporating remedial measures as part of the construction program, many of these problems can be alleviated.

Operation

During the operation phase of the project, the solid waste landfill/disposal sites would be prone to wind and runoff erosion. The application of re-vegetation techniques would help to stabilize the soils and minimize the loss and leaching of soils.

Post-Operation

After decommissioning, the entire exposed area of the project should be re-vegetated. This would minimize the erosion and leaching of soils.

10.2.4 Water Resources

There are two types of water resources in the project area:

i) surface water: Elk River, Fording River, and
ii) groundwater.

10.2.4.1 Surface Water

Construction

Construction activities causing soil erosions may have adverse effects on the water quality of Elk River and Fording River. Disturbed land is prone to wind and runoff erosions, which would re-deposit the soil particles in low lying areas. During periods of rainfall and following snow melt, re-suspension of these particles would increase the concentration of suspended sediments in the surface water to above normal level.

The increased sediment loads in local watercourses may result in increased turbidity, increased oxygen demand and possible siltation of any existing spawning areas. However, a variety of techniques have been developed which, when applied appropriately, can effectively control sediment problems.

A number of other activities associated with the plant construction may also have adverse impacts on the water quality of Elk River or Fording River. These include: in-stream construction activities, withdrawal and discharge of water and construction of access roads. The impacts resulting from these activities are usually considered short-term, local in nature and as being responsive to a variety of relatively simple remedial measures.

Operation

Under normal operating conditions, impacts upon the river waters are expected to be minimal. Liquid and other solid wastes from the process complex would not pose a problem as adequate treatment and storage facilities would be designed to meet all regulatory requirements. Solid residues from the Vacuum Distillation Unit, containing hydrocarbons which have carcinogenic potentials, would be incinerated to eliminate possible seepage problems associated with landfilling.

Storm runoff from the plant site may pick up potentially toxic chemicals which may eventually find their way into the rivers. Properly designed site drainage control system would eliminate this problem.

Sediment loadings of the rivers during the operation phase would not be a significant problem if care is taken in land-filling the solid wastes. Problems may occur when storm runoff carries away landfill waste, thereby increasing the sediment loading in the rivers. Containment system designed as part of the landfill site would eliminate these problems.

The process water make-up requirements may be of some concern from an environmental point of view. Excessive drawdowns of river water and increased wastewater discharges may generate a variety of problems, including: impeding the aquatic life and the water supply source for local communities, gravel bed movement, and bank erosion. To minimize these problems, maximum re-use of the process wastewater would be planned; meanwhile careful withdrawal of water from a combination of sources including rivers, groundwater, and nearby lakes, would be carried out during the operational phase of the project.

10.2.4.2 Groundwater

Construction

Impacts upon groundwater are generally site specific because of regional and local variations in geology and hydrology. The compaction of soil for the plant site with eventual paving of large areas can reduce the amount of groundwater recharge occurring in the area. The construction activity could result in a rearrangement of the piezometric surface of local aquifers in the overburden.

Operation

The use of an impervious liner such as natural clay barrier or synthetic liner, at the landfill site, would prevent any seepage from the solid wastes, thus eliminating its potential impact on the groundwater.

During the operation phase conflicts for water may continue, if water is to be drawn from local aquifers. An overview of the regional geology and the groundwater table would be required in order to obtain an appreciation of the ability of local supplies to satisfy the requirements.

Post-Operation

Post-operational impacts would be limited and seepage from the waste landfill sites would not be a problem if the sites are properly lined and the containment systems are in effective operation.

10.2.5 Atmospheric Environment

Construction

During the construction of the coal liquefaction complex, dust and noise can be expected to have the most significant impacts upon the atmospheric environment. Noise and dust emissions may be caused by vehicle and equipment movement, generators, pumps, compressors and impact equipment such as jack hammers, etc. These impacts tend to be very localized and short-term in nature.

Operation

During the operation of the plant, the principal emissions of concern would be sulphur dioxide, carbon monoxide and particulate solids. Of minor concern would be dust raised by vehicles moving on unpaved roads. These emissions are not beyond known technologies for their control. The process complex would be designed to meet all federal and provincial standards and to minimize all emissions of contaminants.

Depending on the final location of the complex and the process used, the incremental effect of the air emmissions on the ambient air quality is likely to be marginal. Providing that the disposal of solid wastes during the operation of the facility is handled carefully, the ambient air quality in the area should not have any noticeable changes.

It is expected that noise diffused from the process complex would not be excessive. Process equipment generating abnormal noise levels would be separately enclosed, when practical, and sound proofed. This would minimize the noise problems in the plant.

Post-Operation

After the cessation of the processing activities and until the area is re-vegetated, which may take several years, the prevailing winds could erode the surficial layer of soil, thus exposing the waste solids beneath. However, post-operational monitoring of the area while re-vegetation is taking place and the use of known techniques in re-vegetation and reclamation of such areas can avoid this problem.

10.2.6 Terrestrial Environment

Construction

Construction activities would have impacts on the vegetation at the project site, and the wildlife in the vicinity.

Vegetation on the construction site would be largely removed. Wind-blown dust from construction operations would be deposited on the vegetation in the surrounding area, leading to a decrease in photosynthetic activity and a decrease in primary productivity. These effects, although not quantified, would be minor relative to the total loss of vegetation from the site.

Any wildlife residents on the construction site would be displaced or affected. Many small mammals would be destroyed or die rather than simply move to nearby undisturbed areas. This is due to the fact that these small mammals have home ranges that are relatively small. These losses are similar for any industrial or commercial installation.

Other wildlife, including bird populations, in the vicinity may be disturbed by the noise and human activities associated with construction. It is doubtful that there are any rare or endangered species within the project area. However, further investigation will be necessary to ascertain the presence or absence of any wildlife in order to determine the actual impact which this project will have on this component of the environment, if any.

Operation

Impacts upon vegetation are not expected to be great during the operational phase of the plant. Plant emissions of sulphur dioxide and particulate solids are expected to be within the Federal and Provincial standards and, therefore, would not result in damage to terrestrial flora. Vegetation aligning unpaved access roads would be covered in dust due to increased road traffic.

The plant should not cause any significant impact on wildlife. This is because noise-sensitive species would have abandoned the area during the construction phase, and other wildlife would have adapted to the noise and human activities.

Post-Operation

Impacts on the terrestrial environment following completion of project life span would be the reduction of food resources, and the loss of wildlife habitat. Members of higher trophic levels would be least affected since their foraging habits generally encompass a larger area.

10.2.7 Aquatic Environment

Construction

Most impacts on the aquatic environment associated with construction are short-term and local in nature. The largest impact would be caused by increased sediment volumes in the Elk River and Fording River. However, as previously mentioned there are numerous techniques which can be applied, e.g. sediment trap, hay bales, coffer dams, etc. to reduce sediment loadings to a tolerable level.

There are several ways in which increased sediment input to a waterbody may affect freshwater biota. The impacts most commonly associated with excess sediment are: habitat disruption by removal of sediments, reduction in light penetration due to increased quantities of suspended soil particles, abrasive effects of suspended sediments on sessile organisms, sediment dispersion, fall-out and the smothering of benthic invertebrates (bottom dwelling organisms), aquatic insects and fish eggs and changes in water quality that could either stimulate or inhibit biological growth.

Benthic communities would be adversely affected by increased sedimentation. The recovery rate of these communities depends on the physical, chemical, and biological characteristics of the sediment addition and the availability of recolonizing organisms. Estimates for the recovery of benthic communities distrubed by excavation range from 2 to 15 months. These impacts are not considered significant because they are of low magnitude and short duration. There would be some unavoidable impacts but of a localized nature.

Increased amounts of sediment within the water may also affect fish eggs. Fish eggs are susceptible to smothering when suspended sediments are deposited on spawning beds. Short-term, localized increases in sediment deposition may be tolerable but the effect of long-term increases could result in the loss of spawning areas.

Adult fish may find fewer benthic organisms and aquatic insects available due to increased sedimentation. Since this impact would be of a localized nature, it is not expected to have any significant effect on fish productivity.

Operation

Impacts upon the aquatic environment are not expected to be great during the operation of the facility. Potential problems could arise if improper treatment of solid and liquid wastes occurs. The taking of process and potable water from the water source could cause problems within the aquatic community. The possibility of draw-down during low-flow months would adversely impact on the aquatic organisms present. The water intake pipe should not be located near any spawning areas to avoid affecting fish productivity locally.

10.3 CONTROL MEASURES

10.3.1 General

In consideration of the potential impacts on the environment, appropriate mitigation measures would be implemented during the construction, operation and post-operation phases of the project. These include: construction monitoring, erosion/siltation controls, site drainage control, wastewater treatment, air emissions control, solid wastes management, re-vegetation and slope stabilization.

The proposed environmental control programs/systems are designed on the basis of best available control technology to ensure the following British Columbia regulatory requirements are satisfied:

- Pollution Control Act	R.S.B.C. 1979, C.322
- Pollution Control Regulation	B.C. Reg. 96/67
- Pollution Control Objectives for the Chemical and Petroleum Industries of British Columbia	1974
- Health Act	R.S.B.C. 1979, C.161
- Sewage Disposal Regulations	B.C. Reg. 577/75

-	Minimum Requirements for Refuse Disposal to Land	1971
-	Soil Conservation Act Regulations	B.C. Reg. 77/74

In addition, necessary in-plant health and safety protection measures would be implemented to ensure a satisfactory working environment for the plant workers.

10.3.2 Construction Monitoring

During construction, earth removal and equipment operations would be monitored and the following precautions would be taken to ensure that the disturbance to the environment is reduced to a minimum:

(a) equipment or vehicle operations would be carried out carefully so as not to make any unnecessary disturbance to or removal of vegetation cover;

(b) in the area where the vegetation cover is removed, particularly on the disturbed slopes, ditches would be constructed to intercept and divert the surface runoff, therefore reducing the erosion effects;

(c) sediment traps would be installed, wherever applicable, to prevent eroded soils entering the water streams;

(d) when practical, re-vegetation of disturbed land and stabilization of disturbed slopes would be carried out as early as possible to minimize wind and runoff erosions.

10.3.3 Water Supply

A total fo 18 000 m^3/d of raw water make-up is required for the process complex.

Pending the detailed investigation on the availability of water sources, raw water would be drawn from one or more of the following water bodies: Elk River, Fording River, groundwater or nearby lakes. Caution would be taken not to impede any water supply for local communities and the aquatic life dependent upon the streams.

The alternative would be to maximize the re-use of process wastewater, in order to reduce the raw water make-up requirements. The possibility of treating and re-use of cooling tower blowdown and neutralized boiler feedwater plant effluent would be assessed in the detailed engineering phase.

10.3.4 Wastewater Treatment and Disposal

10.3.4.1 Process Foul Water

Foul water from the H-Coal Liquefaction Units, Hydrotreaters and Residue Gasification Units contain high concentrations of phenolic compounds, ammonia, carbon dioxide and traces of hydrogen sulphide.

The combined process foul water is sent to a Phenol Recovery Plant, from which the recovered phenols are pumped to a Fuel System and burned as process fuels. The remaining liquid in the Phenol Recovery Plant is sent to an Ammonia Recovery Plant, in which the ammonia and hydrogen sulphide levels of the liquid are reduced significantly; the recovered ammonia is stored, shipped and sold as anhydrous ammonia.

The effluent from the Ammonia Recovery Plant contains minor quantities of phenols, ammonia, carbon dioxide and hydrogen sulphide. This effluent, together with other wastewaters, are sent to a Biological Treatment Plant for further treatment to within acceptable quality levels, prior to their discharge to the environment.

10.3.4.2 Other Wastewaters

Other wastewaters received by the Biological Treatment Plant include sewage generated in the process complex, blowdown from the cooling tower and neutralized regeneration effluent from the boiler feedwater treatment plant. The plant sewage is of domestic type; the cooling tower blowdown contains suspended solids and moderate hardness; the neutralized boiler feedwater plant effluent contains salts, mainly sodium sulphate.

10.3.4.3 Biological Treatment Plant

Prior to its discharge into receiving water bodies, a total of 140 t/h (or 140 m^3/h) of wastewater is treated in an "activated sludge" treatment plant consisting of primary clarifier, aeration tanks, secondary clarifier and chlorination unit. The quality of treated effluent is maintained within the limits set by the Pollution Control Objectives for the Chemical and Petroleum Industries of British Columbia: $B.O.D._5$, 20 mg/L; suspended solids, 20 mg/L; phenols, 0.2 mg/L; ammonia, 10 mg/L, sulphides, 0.10 mg/L.

10.3.5 Site Drainage Control

10.3.5.1 Plant Site Runoff

The site runoff is intercepted by the ditches around the process complex, and flows by gravity to a collection pond. The uncontaminated runoff, when necessary, is discharged to a suitable water body for disposal.

10.3.5.2 Emergency Coal Stockpile Drainage

The drainage from the emergency coal stockpile has a low pH. It is intercepted by the ditches around the stockpile, and flows by gravity to a collection pond. In the collection pond the drainage is treated by 10% $Ca(OH)_2$ to pH 7.0. When necessary, the neutralized drainage is discharged to a suitable water body for disposal.

10.3.5.3 Chemical Reagent Spills

Solid chemical spills would be recovered in the solid state; the recovered spills would be dissolved in the solution preparation tank for process use.

Solution spills in the plant areas would be "squeegeed" to the sump and the floor washed with water. To facilitate the spill collection, washdown and clean-up operations, the floor of the process plant areas are sloped and collection sumps are provided wherever necessary. The recovered spills would be re-used, where practical, or treated with neutralizing agents prior to disposal.

In the event of chemical spills outside the plant areas, contaminated soils would be treated with appropriate neutralizing agents and disposed of in a designated landfill site. When absorbents, such as peat moss are used, the reclaimed spills would be incinerated in a controlled manner.

All of these measures would ensure adequate containment and disposal of chemical spills.

10.3.6 Air Emissions Control

Air emissions from the process complex to the atmosphere include gaseous effluents from the Oxygen Plant, the Residue Gasification units, the Light-Ends Recovery Plant, the Fuel Gas System, the Sulphur Recovery System, and the Boiler Plant. The concerned quality parameters are: sulphur dioxide, carbon monoxide and particulate solids.

In general, the method of air emissions control is to collect the air exhaust from the source equipment, recover or remove the air contaminants by best available control technology devices, and discharge the cleaned air exhaust via a stack where further dilution of air emissions would occur.

The details of air emission control measures are given in the following sub-sections.

10.3.6.1 Sulphur Dioxide

Sulphur in the raw coal (0.3% wt) is converted to hydrogen sulphide in the H-Coal Liquefaction Units, or remains with the raw naphtha, middle and heavy distillate products, or the solid residues. Sulphur compounds in the raw naphtha and middle distillate products are converted to hydrogen sulphide in the Naphtha and Distillate Hydrotreaters.

The hydrogen sulphide produced in the H-Coal Liquefaction Units and Hydrotreaters is contained in the process fuel gas produced by these units. The main fuel gas stream from these units is treated in a Sulphur Plant where hydrogen sulphide is converted to elemental sulphur. Hydrogen sulphide removal efficiency in the Sulphur Plant is 99% minimum, resulting in the production of a sweet fuel gas containing less than 15 mg/m^3 hydrogen sulphide. The combusted fuel exhaust contains less than 30 mg/m^3 sulphur dixoide, well below the limit (830 mg/m^3) set by the British Columbia Pollution Control Objectives.

The solid residue, containing about 1.0 wt% sulphur, is used as the fuel in the Boiler Plant for the generation of electricity and process steam. Sulphur dioxide in the boiler flue gas is recovered in a Flue Gas Desulphurization Unit, and disposed of as solid gypsum and other forms of calcium sulphate. The treated flue gas contains a sulphur dioxide concentration well below the permissible limit, 830 mg/m^3.

10.3.6.2 Particulate Solids

The major emission source of particulate solids is the Boiler Plant which uses the solid residue from the Vacuum Distillation Unit as fuel. The Flue Gas Desulphurization Unit is designed to produce a flue gas containing less than 100 mg/m^3 particulate solids (B.C. regulatory limit is 150 mg/m^3).

Throughout the process complex, dust collection and particulate removal systems are provided wherever the potential of dust or particulate solids generation exists. Pneumatic unloading, transfer and feed systems are utilized to handle the solid catalysts in the H-Coal Liquefaction Units and the lime in the Flue Gas Desulphurization Units. Cyclone separators and dust collection system are provided at each transfer point to minimize particulate emissions. The dust collection systems are designed to reduce the particulate solids level to less than 150 mg/m³.

Precautions have been taken in the Coal Handling Area to minimize dust emissions. Both the coal stockpile and the residue stockpile are enclosed in buildings which are equipped with dust collection facilities; all outside coal conveyors are also enclosed in galleries.

All of these dust control measures would help to provide a safe working environment for the workers.

10.3.6.3 Carbon Monoxide

The carbon monoxide concentrations of gaseous effluents from the Residue Gasification Units, the Fuel System and other plant units, are all well below the limit set by the British Columbia Pollution Control Objectives, 2 400 mg/m³.

10.3.6.4 Flare Systems

Smokeless flare systems are provided to accept relief and vent vapours from all process units:

H-Coal Liquefaction Units

A separate dedicated flare system is provided for each H-Coal Liquefaction Unit, each sized for the emergency shutdown and depressurization of one coal liquefaction train. All vapours released during the emergency depressurization are flared and the coal slurry accummulated in the blowdown drum is recycled to the coal slurry feed tank.

Hydrogen Plant

A single flare system is provided for the Hydrogen Plant. It is sized for the emergency shutdown and depressurization of the unit.

Other Process Units

A general plant flare system is provided to accept relief and vent vapours from the Atmospheric and Vacuum Distillation Units, the Hydrotreaters, the Light Ends Recovery Unit, the Sulphur Plant, offsites and utilities. The flare is designed on the basis of maximum overhead vapour rate from the Distillation Units.

10.3.7 In-Plant Health and Safety

In addition to the air emission control facilities described in Section 10.3.6, the process complex is provided with a controlled ventilation system to supply sufficient make-up fresh air to the working areas.

Blowers, vacuum pumps, compressors, ejectors and other process equipment generating abnormal noise levels are separately enclosed, wherever practical, and sound proofed.

All of these measures would ensure a safe working environment for the plant workers.

10.3.8 Solid Wastes Management

A total of 2 240 tonne/day of solid wastes are generated in the process complex. These solids include the solid residues from the Vacuum Distillation Unit, the slags from the Residue Gasification Units, the fly ash and bottom ash from the Boiler Plant, the sludges from the Flue Gas Desulphurization Unit, and the waste sludges from the Biological Wastewater Treatment Plant.

Solid residues from the Vacuum Distillation Unit, 750 tonne/day, are incinerated. This would eliminate possible environmental problems caused by landfilling -- seepage of carcinogenic hydrocarbons into the nearby water bodies. Other collected solid wastes are transported to suitable sites for landfill. Lining and containment structure would be provided, wherever necessary, to prevent any leaching of solid wastes and subsequent seepage into the groundwater or nearby streams.

Alternatively, depending upon the waste characteristics to be confirmed by laboratory tests, some solid wastes may have to be stored in a series of lined lagoons. These lagoons can be built in stages at 5-year intervals.

Volume

2

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Western Canada Coal Liquefaction Study
Prefeasibility Study

Prepared for:

British Columbia Resources Investment Corporation
Petro-Canada Exploration Incorporated
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British Columbia Coal Limited
Energy, Mines and Resources Canada

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Kilborn and Badger have followed standard professional procedures in preparing the contents of this Report. It is based in part on details, information and assumptions provided by others and Kilborn and Badger, therefore, cannot guarantee the correctness of the Report; but to the extent of their investigation and within the scope of the work delegated to them, Kilborn and Badger believe the Report is substantially correct.

Introduction

1.0 INTRODUCTION

1.1 GENERAL

This volume presents the economic aspects of the conceptual coal liquefaction complex, based on an assumed net distillate yield of 40%, maf coal. Results of the economic and sensitivity analysis is presented as well. All costs are in 1982 Canadian dollars for a complex located in south-eastern British Columbia.

1.1.1 Capital Cost Estimates

Capital cost estimates were prepared for a base case and five alternative cases as follows:

Base Case: Incorporating all facilities described in the Technical volume of this report to produce a nominal 24 000 barrels per day of syncrude. Solid residue is used to generate hydrogen and inplant power. Excess residue is discarded at an assumed disposal cost of $10.00 per tonne.

Cost $2.02 Billion.

Alternative Cases

Case 1: Incorporating all Base Case facilities described above to give an assumed 50 percent distillate yield from the liquefaction unit and a nominal 28 500 barrels per day of syncrude.

Cost $2.062 Billion

Case 2: Base Case process facilities with no power generation. Excess residue and fuel gas are gasified and converted via the MTG process to finished gasoline. Electrical power is purchased. Plant capacity is a nominal 24 000 barrels per day syncrude and 8 000 barrels per day gasoline.

Cost $2.520 Billion.

Case 3: Base Case process facilities with no power generation. Coal tailings, excess residue and fuel gas are gasified and converted via the MTG process to finished gasoline. Electrical power is purchased. Plant capacity is a nominal 24 000 barrels per day syncrude and 10 000 barrels per day gasoline.

Cost $2.630 Billion.

Case 4: Base Case process facilities with no power generation. Hydrogen make-up is by steam-methane reforming. Excess residue is gasified and converted via the MTG process to finished gasoline. Electrical power is purchased. Plant capacity is a nominal 24 000 barrels per day syncrude and 17 000 barrels per day gasoline.

Cost $3.039 Billion.

Case 5: Base Case process facilities with inplant power generation. Hydrogen make-up is by steam-methane reforming. Excess residue, not used for inplant steam and power generation, is discarded at zero value or zero disposal cost (on the assumption that it can be utilized elsewhere as boiler fuel or some other type of fuel). Plant capacity is a nominal 24 000 barrels per day syncrude.

Cost $2.087 Billion.

Details of the capital cost estimate and criteria are provided in Sections 1.3 and 2.0 of this volume.

1.1.2 Operating Cost Estimates and Revenue

Operating cost estimates were developed for the base case and five alternatives. Operating costs calculated in 1982 dollars were based on:

- coal requirements, at $80.00 per tonne
- raw material requirements
- utility requirements (alternative cases only)
- manpower estimates for complex staff
- maintenance parts
- taxes and insurance.

Product revenues were also developed for the base case and alternatives based on 1982 dollars. Criteria for product revenues were:

- syncrude based on September 1981 Oil Pricing Agreement
- propane based on September 1981 Oil Pricing Agreement
- butane based on September 1981 Oil Pricing Agreement
- gasoline based on September 1981 Oil Pricing Agreement
- sulphur based on prevailing product prices
- ammonia based on prevailing product prices
- soda ash based on prevailing product prices.

A summary of annual revenue and operating costs in 1982 dollars is provided below.

	$1 000		
	Revenue	Operating Cost	Gross Margin
Base Case	437 500	393 080	44 420
Alternative Cases			
Case 1	494 050	391 725	102 325
Case 2	562 710	452 670	110 040
Case 3	599 180	472 440	126 740
Case 4	677 870	584 640	93 230
Case 5	420 113	494 050	Negative

Details of the operating cost estimate and product revenues are provided in Section 2.0 of this volume.

1.1.3 Economic Analysis

The economic analysis is based upon 100 percent equity financing. Capital expenditures are spread over a seven year period with a three year startup and commissioning period prior to attaining design capacity. Working capital, escalation, capital cost allowances and investment tax credits have been considered in developing the cash flow model.

Using the cash flow model, the DCF rate of return for the Base Case and Alternatives is as follows:

	DCF rate of return (percent)	
	$80/tonne coal	$50/tonne coal
Base Case	4.6	10.2
Alternative Cases		
Case 1	8.4	12.7 (Cal.)
Case 2	7.6	
Case 3	8.1	
Case 4	7.7	
Case 5	Negative	

Sensitivity analyses were run on the Base Case at two coal prices to test the effect of variations of capital cost, product revenue and operating cost on the DCF rate of return. These are given in Section 3.0 of this Volume.

Three special sensitivity analysis cases were run, as summarized below with their respective DCF rates of return:

	DCF rate of return (percent)
Base Case with corporate tax calculation, rather than single project basis	6.6
Base Case with $10.00 per barrel syncrude discount	Negative
Base Case with $60.00 per tonne coal, $5.00 per barrel syncrude discount and adjusted escalation factors	9.6

1.1.4 Addendum to Economic Analysis

Project DCF rates of return were calculated for the Base Case and Alternative Cases 1-5 based on an April 13, 1982 Announcement by the Province of Alberta, in which a revised forecast of new oil reference prices was presented to reflect recent changes in projected World Oil prices. Based on this Announcement, the H-Coal syncrude price was calculated to be $56.85 per barrel, effective July 1986, compared to $72.40 per barrel using the September 1981 Agreement pricing structure.

Project DCF rates of return are as follows for the Base Case and Alternatives, at the two coal prices:

	DCF rate of return (percent)	
	$80/tonne coal	$50/tonne coal
Base Case	Negative	4.3
Alternative Cases		
Case 1	0.0	7.7
Case 2	Negative	6.1
Case 3	Negative	6.4
Case 4	Negative	4.8
Case 5	Negative	Negative

This analysis clearly indicates that the effect of the lower product prices is to reduce DCF rates of return substantially at equivalent coal prices. In fact, indicated DCF's are negative or zero for $80.00 per tonne coal and coal prices must be reduced by $30.00-$40.00 per tonne to achieve DCF rates of return, similar to those obtained in the original analysis, with the lower product prices.

1.2 BASIS FOR ECONOMIC ANALYSIS

1.2.1 General

The economic and sensitivity analysis for the coal liquefaction complex was prepared to determine the economic feasibility of the project and to study various plant configurations. The economic model is a cash flow generator for before and after tax calculations. Developed within this model are certain assumptions used for this analysis. The model criteria and data resources are briefly described in this section.

1.2.2 Reference Date and Project Life

This project has been evaluated to a January 1, 1982 reference date and assumes a 27 year project life. After a three year period for detailed feasibility studies and PDU coal testing, a four year construction period is scheduled followed by a 20 year production profile.

1.2.3 Yearly Inflation

The yearly inflation factors used in this study are based on the GNP Deflator and Consumer Price Index Forecasts provided in the Conference Board 1981 Mid-term Forecast shown in Table 1.1. An averaging of these inflation factors resulted in an average inflation rate of nine percent up to 1984 and seven percent for 1985 and up.

Capital by-product values and operating costs are escalated a further two percent over the project life. Liquid hydrocarbon product values are escalated to 1986 based on the September 1981 oil agreement and beyond 1986 based on a combined rate of nine percent (seven percent inflation and two percent real escalation). Debt required to finance working capital or operating costs is serviced with an interest rate equal to the above inflation structure plus four and one-half percent.

1.2.4 Construction Schedule

Capital costs as estimated in 1982 dollars were distributed by percentage to the year of construction following the construction schedule outlined in Figure 1.1 and escalated to the appropriate year. The schedule for capital expenditures is as follows:

Year	Percent Capital	Percent Completion
1982	Nil	Nil
1983	Nil	Nil
1984	5	5
1985	15	20
1986	25	45
1987	30	75
1988	25	100

1.2.5 Operating Costs

The design operating costs based on feedstock requirements, catalyst and chemical requirements, maintenance, manpower, insurance and taxes are presented in Section 2.0 of this Volume. In all cases evaluated the raw material requirements for coal represented the major operating expenses in the operating cost estimates.

All operating costs were calculated in 1982 and escalated to the relevant year using the inflation factors provided.

1.2.6 Plant Production

Plant production is expected to be phased in as per the startup and construction schedule in Figure 1.2. Plant capacity in the first year of operation is expected to be about 40 percent with the commissioning of three of the five H-Coal liquefaction trains. Plant capacity is increased in the second and third years of operation to 70 and 90 percent of design capacity with the commissioning of the fourth and fifth H-Coal trains. Full capacity is attained in the fourth year of operations and maintained for the duration of the plant life.

1.2.7 Product Prices

The following 1982 product prices were used for products from the coal liquefaction complex:

Product	1982 Product Price
Propane	$ 42.00 per barrel
Butane	42.00 per barrel
Syncrude	42.90 per barrel
Stabilized Gasoline*	50.00 per barrel
Sulphur	100.00 per tonne
Ammonia	260.00 per tonne
Soda Ash	65.00 per tonne

* alternative cases only.

Liquid hydrocarbon prices were escalated according to the September 1981 Oil Pricing Agreement in Table 1.2 to 1986. Prices are escalated at nine percent from 1987 on. By-product prices presented in 1982 are escalated according to the inflation factors described in Section 1.2.3 of this Volume.

1.2.8 Taxation

As per current federal regulations, income from this complex was considered manufacturing and processing profits. Taxable income is calculated as revenue less operating costs less interest paid on outstanding debt less capital cost allowances. Capital cost allowances are divided into two categories based on the end use of the expenditure. Process related expenditures representing an estimated 98 percent of the fixed capital investment are designated as Class 29 for fast write-off purposes. The remaining capital investment is assumed to represent plant buildings and therefore designated as Class 3.

Class 29 expenditures are assigned a maximum capital cost allowance of 25 percent in the first year of operation, 50 percent in year two and 25 percent in year three. Any allowances not fully utilized in any particular year are carried forward to reduce the following year's taxable income. For example, if only 10 percent of the allowance is needed in year one to reduce taxable income to zero, a 65 percent allowance is available in year two. Class 3 expenditures are assigned an allowance of 5 percent per year, declining balance.

Investment tax credits of 10 percent were applied to eligible taxable income. Depletion and resource allowances were assumed to be ineligible for use in this evaluation.

Federal income tax and British Columbia income tax are calculated as 36 percent and 9 percent of taxable income, respectively.

TABLE 1.1

YEARLY INFLATION

(percent)

YEAR	GNP DEFLATOR	CONSUMER PRICE INDEX
1981	9.5	12.6
1982	10.3	10.5
1983	9.1	8.8
1984	8.1	7.9
1985	6.8	6.6
1986 to 2008	6.9	6.9

* Source: Conference Board 1981 Mid-term Forecast

TABLE 1.2

PRICING STRUCTURE FOR H-COAL HYDROCARBON PRODUCTS

			(dollars/barrel)		
DATE	NEW OIL REFERENCE PRICE	H-COAL SYNCRUDE PRICE	PROPANE	BUTANE	GASOLINE
Jan. 1982	45.92	42.90	42.00	42.00	50.00
July 1982	49.22	46.00	45.00	45.00	53.60
Jan. 1983	53.06	49.60	48.50	48.50	57.80
July 1983	57.06	53.30	52.20	52.20	62.20
Jan. 1984	60.18	56.20	55.00	55.00	65.60
July 1984	63.48	59.30	58.10	58.10	69.20
Jan. 1985	66.83	62.40	61.10	61.10	72.80
July 1985	70.23	65.60	64.20	64.20	76.50
Jan. 1986	74.08	69.20	65.80	65.80	80.70
July 1986	77.48	72.40	70.90	70.90	84.40

REFERENCE: September 1981 Oil Pricing Agreement

FIGURE 1.1

ENGINEERING & CONSTRUCTION SCHEDULE

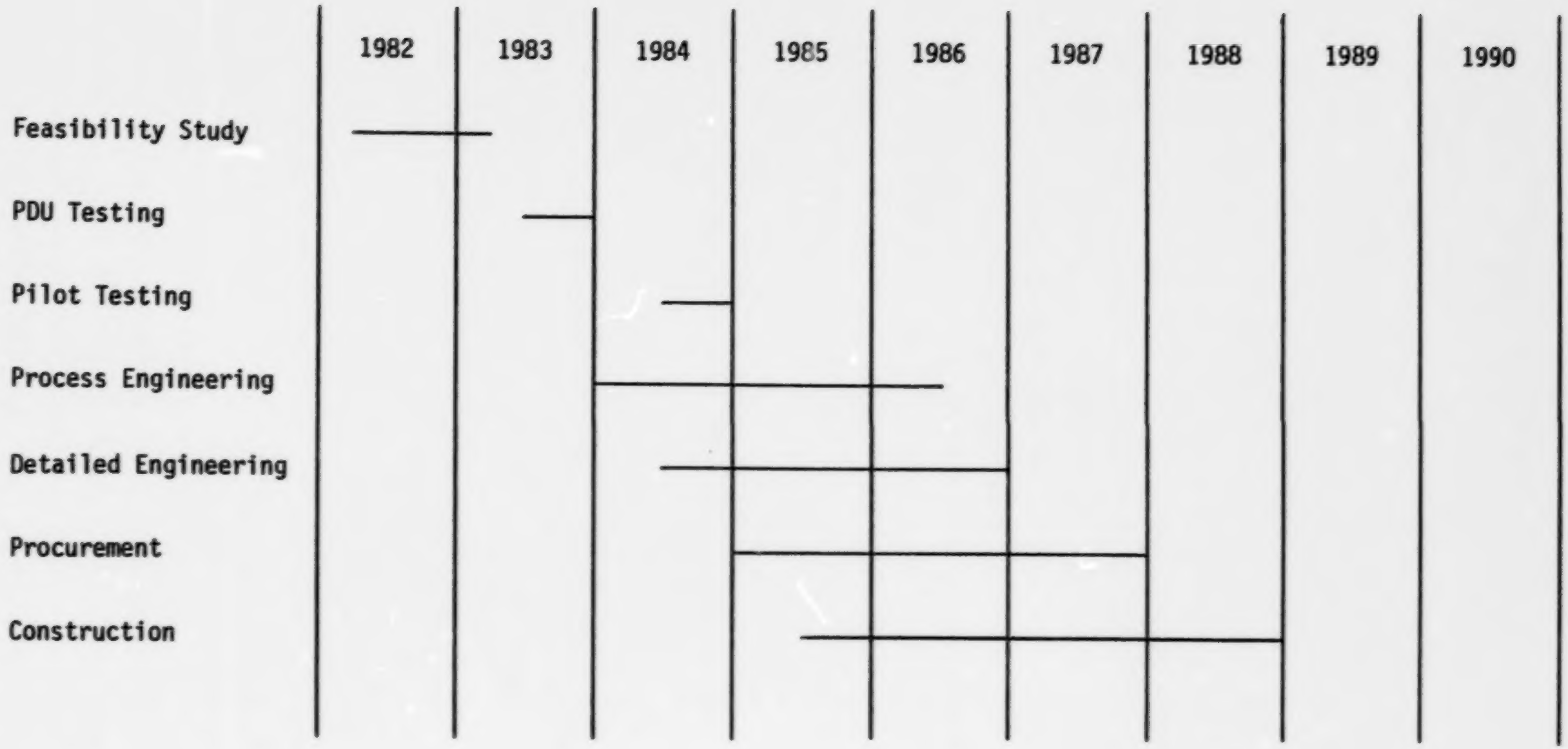
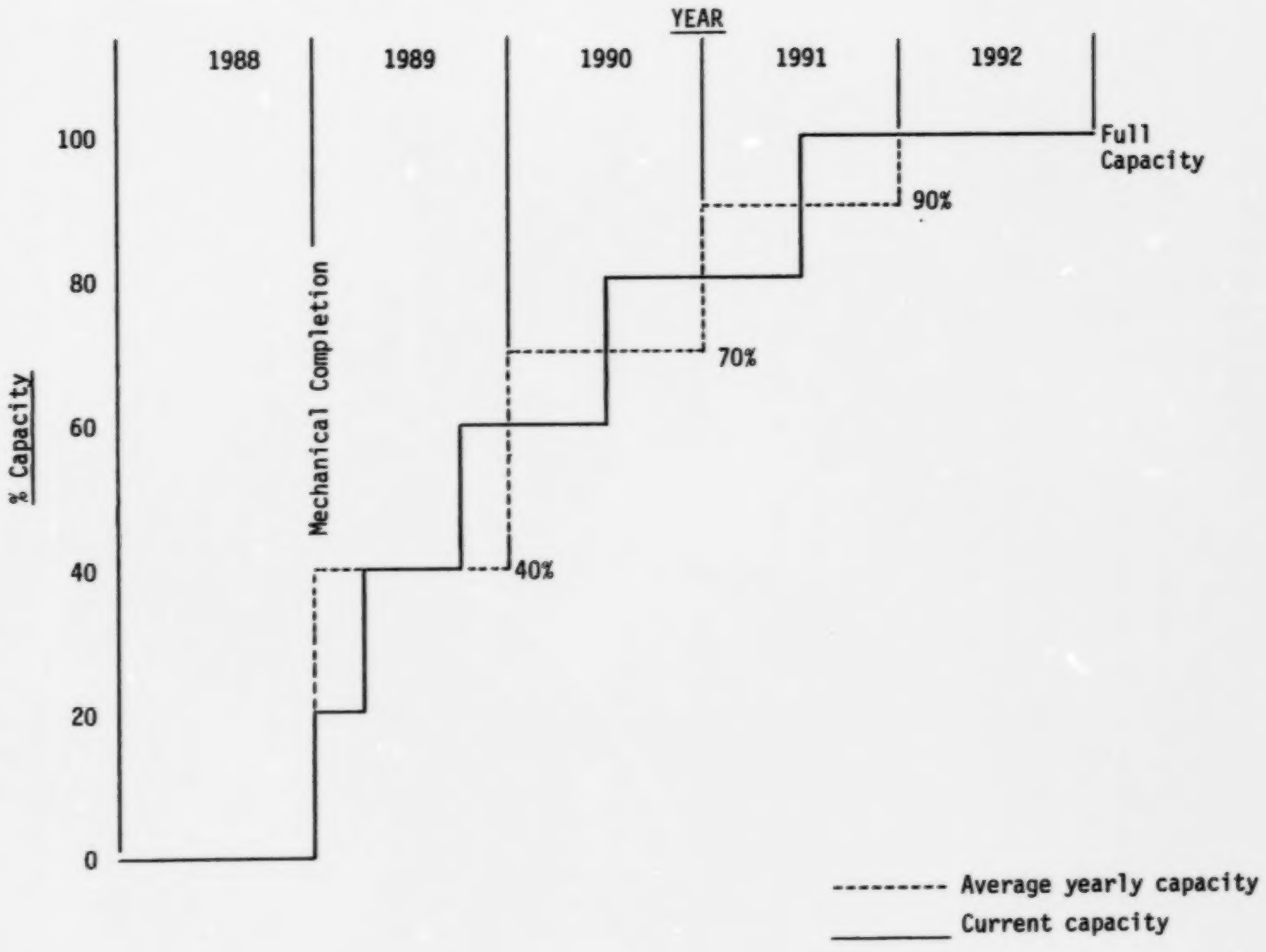


FIGURE 1.2

STARTUP & PRODUCTION SCHEDULE



1.3 BASIS FOR CAPITAL COST

Capital cost estimates are provided for six cases, a Base Case and five alternative cases. All costs are provided in first quarter 1982 Canadian dollars.

Section 2.1 presents the capital cost estimated for the Base Case coal liquefaction complex. This is the case described throughout the report and includes all the facilities described in Section 7.0 and 8.0 of Volume I. The process equipment estimate summary (Section 2.7) contains the back-up information used to prepare the Base Case capital cost. The equipment price summary (Section 2.8) details the capital cost of equipment required in Area 300 - Coal liquefaction and Area 400 - Atmospheric & Vacuum Distillation.

Section 2.2 presents the capital cost summaries for the five alternative cases:

Case 1: 50 percent distillate yields;

Case 2: Fuel gas and residue partial oxidation to MTG;

Case 3: Fuel gas, residue and coal wash plant tailings partial oxidation to MTG;

Case 4: Steam-methane reforming to hydrogen, residue and coal wash plant tailings partial oxidation to MTG.

Case 5: Steam-methane reforming to hydrogen, residue to steam and power generation.

In Case 1, the distillate yield from the coal liquefaction unit is increased to 50 percent resulting in complete utilization of all residue and fuel gas to supply the plant steam, power and hydrogen requirements. Case 2 presents an optimization of liquid products with electrical power purchased and excess residue and fuel gas directed to gasoline production. Case 3 is an extension of Case 2 wherein coal tailings from an offsite beneficiation plant are added to the excess residue and fuel gas for gasoline production. Case 4 is a variation of Cases 2 and 3 where natural gas is purchased for hydrogen production and residue and coal washing tailings are converted to gasoline. Case 5 is a variation of the Base Case wherein natural gas is used for hydrogen production and all residue not used for inplant steam or power generation is discarded.

The Base Case capital cost estimate is based on cost data developed for each process unit, offsite facilities and utility systems. Included in the process units and offsite facilities are proprietary and non-proprietary processes. Cost information from process vendors was used in developing process equipment direct costs for the following process packages:

Naphtha Hydrotreating
Middle Distillate Hydrotreating
Heavy Distillate Hydrotreating
Cryogenic & Pressure Swing Adsorption Hydrogen Recovery
Stretford Sulphur Plant
Steam Generation
Power Generation
Flue Gas Desulphurization
Phenol Recovery
Ammonia Recovery.

Economic data for the above process packages except steam and power generation were provided in U.S. dollars, Gulf Coast location. These data were converted to Canadian dollars, Greenhills location using the factor:

Canadian dollars, Greenhills location = 1.44 x U.S. dollars,
Gulf Coast location.

The 1.44 conversion factor is made up of two components -- a materials component representing 70 percent of the value of the plant and a labour component representing 30 percent of the value of the plant.

For the materials component, it is assumed that 60 percent of the equipment and material will be purchased domestically and 40 percent will be purchased internationally, and therefore subject to a 17 percent import duty. A monetary exchange factor of 1.25 was used to convert U.S. dollars to Canadian dollars.

Materials Factor = 0.7 (0.6 x 1.25 + 0.4 x 1.25 x 1.17) = 0.94

For the labour component construction workers in Canada are assumed to earn the same hourly wages in Canadian dollars as workers in the Gulf Coast earn in U.S. dollars. Labour productivity multiplier in the Greenhills location is assumed to be 1.40 compared to the Gulf Coast. A labour premium factor of 1.20 was used as recognition of the need to increase labour wages to attract workers to the plant site.

Labour Factor = 0.3 (1.20 x 1.40) = 0.50

Conversion Factor = Materials Factor + Labour Factor
= 0.94 +0.50 = 1.44

Cost information for steam generation was provided in U.S. dollars, Gulf Coast location but, unlike other process packages, did not include auxiliary equipment. A factor of 1.9 was used to convert U.S. dollars, Gulf Coast location to Canadian dollars, Greenhills location, while taking cognizance of the required auxiliary equipment.

All other process packages were costed in U.S. dollars, Gulf Coast location using Kilborn in-house information. These data were converted to Canadian dollars, Greenhills location using the 1.44 factor. Costs for Area 300 - Coal Liquefaction and Area 400 - Atmospheric and Vacuum Distillation process equipment, other miscellaneous process unit equipment, offsite facilities and utilities were developed using Kilborn in-house estimating techniques for budget pricing of major equipment items. Equipment prices provided in U.S. dollars Gulf Coast location were converted to Canadian dollars, Greenhills location using the factor:

Canadian dollars, Greenhills location = 1.33 x U.S. dollars, Gulf Coast location.

The 1.33 factor is developed assuming 60 percent of the equipment and material is purchased domestically and 40 percent is purchased internationally, and therefore subject to a foreign equipment import duty of 17 percent. A monetary exchange factor of 1.25 was used to convert U.S. dollars to Canadian dollars.

$$\text{Factor} = (0.6 \times 1.25) + (0.4 \times 1.25 \times 1.17) = \underline{1.33}$$

Factors were applied by Kilborn for equipment transportation, installation, process piping, electrical, structural and instrumentation to give process equipment direct cost. Factors were applied to all process equipment direct costs for interbattery piping, site development, buildings, and site access to give total direct costs.

In summary, factors used to develop total direct plant costs are as follows:

Major equipment cost	+ 5-10% transportation + 8-12% for equipment installation

to give total process equipment installed cost:

+ 60% for process piping
+ 20% for instrumentation
+ 15% for electrical
+ 15% for process structures

to give total direct process equipment:

+ 5% for interbattery piping
+ 12% for site development
+ 2% for buildings
+ 0.2% for site access

to give total direct costs:

+ 35% of above for engineering
  and construction
+ 25% of above as contingency

to give total project cost.

The contingency used in the preparation of the capital cost estimate is 25 percent. This contingency is actually an average based on the degree of process uncertainty and the degree of project uncertainty in the cost estimates of the individual plant units. The process and project contingencies used to develop the average contingency are listed below.

Area		Contingency	
		Process	Project
200	Coal Preparation & Handling	Nil	20
300	Coal Liquefaction	15	20
400	Atmospheric & Vacuum Distillation	5	20

Area		Contingency	
		Process	Project
500	Naphtha Hydrotreating	10	20
600	Middle Distillate Hydrotreating	10	20
700	Heavy Distillate Hydrotreating	10	20
800	Light Ends Recovery	Nil	20
900	Hydrogen Plant	5	20
1000	Oxygen Plant	Nil	20
1100	Sulphur Plant	Nil	20
1200	Steam & Power Generation	Nil	10
1300	Phenol Recovery	Nil	20
1400	Ammonia Recovery	Nil	20
1500	Offsites	Nil	20
1600	Utilities	Nil	20

The Kilborn capital cost estimate represents an accuracy of plus or minus 30 percent or one chance in three that the actual plant cost will exceed the estimated cost by greater than 10 percent. No allowance has been made for "outside battery limits" infrastructure costs that could conceivably add to the overall project cost. These excluded infrastructure costs include such items as:

- housing and community services for permanent operating staff
- power transmission to the site (alternative cases)
- coal receiving facilities beyond the plant battery limits
- disposal of residue and ash from the site
- land costs and other miscellaneous owner's costs.
- natural gas transmission to the site (alternative cases)

1.4 Basis for Operating Cost Estimate

Operating cost estimates were prepared for six cases (a Base Case and five alternative cases), based on an operating year of 310 days and consists of direct production costs (raw materials, manpower, utilities and maintenance) and fixed overhead charges (local taxes and insurance). All costs are provided in first quarter 1982 Canadian dollars.

Section 2.3 presents the details of the operating cost estimate for the Base Case coal liquefaction complex. Operating cost summaries for the alternative cases are provided in Section 2.4.

Overall manpower for the complex is 860, based on the operating staff plan as given in Volume I of this report. Current labour rates for western operating facilities are \$12.34 per hour for labourers and \$15.95 per hour for tradesmen excluding all fringe benefits. A wage rate for the complex was developed by a weighted averaging of the above rates as follows:

$$(\$12.34 \times \frac{2}{3}) + (15.95 \times \frac{1}{3}) = \$13.54 \text{ per hour}$$

Fringe benefits were estimated to be 35 percent of the base rate resulting in a final wage rate of \$18.28 per hour or \$38 000 per year.

Coal mined at the adjacent mine at a strip ratio of 10:1 is beneficiated to reduce the ash and water content. Beneficiated coal is valued at \$80.00 per tonne.

The Base Case complex is self-sufficient in electrical power and fuel gas and pumps its own water for internal use. However, electrical power and natural gas are required in selected alternative cases. In this cases the cost of electrical power and natural gas is determined using a cost per unit basis as follows:

Electricity	\$ 0.0226 per kilowatt hour
Natural gas	\$ 0.149 per normal cubic metre

Catalyst and chemical quantities and costs are based on process licensor information and current Canadian chemical prices. Coal liquefaction catalyst consumption is estimated to be three times greater than typical consumption rates for Illinois No. 6 coal.

Parts required for maintenance are estimated at one percent of the total plant cost while maintenance labour costs are included in the overall plant manpower staffing. An allowance of $2 000 000 per year has been provided for emergency subcontracted maintenance.

Local taxes and insurance are estimated to be two percent of the total plant cost plus working capital per year.

1.5 BASIS FOR PRODUCT PRICING

Liquid hydrocarbon pricing for the coal liquefaction complex is based on the September 1981 oil agreement wherein a new oil reference price of $45.90 per barrel was established, effective January 1982. H-Coal naphtha, middle distillate and heavy distillate is blended as a synthetic crude for transportation and pricing purposes. This syncrude is priced according to the new oil reference price of $45.90 per barrel less a hydrogen deficiency penalty of $3.00 per barrel. The hydrogen deficiency penalty represents an estimated cost of $1.00 per 1 000 standard cubic feet of hydrogen required to upgrade H-Coal syncrude to an oil sands quality syncrude.

Stabilized gasoline produced in Alternative Cases 2, 3 and 4 is valued at $50.00 per barrel as of January 1982. Cognizance of the finished quality of this product relative to syncrude was taken when establishing a price.

Propane and butane as LPG was discounted to a value of $42.00 per barrel based on their energy content relative to other finished products.

By-products from the complex, including elemental sulphur, anhydrous ammonia and 57% wt soda ash, were priced at current 1982 values of $100.00 per tonne, $260.00 per tonne and $65.00 per tonne, respectively.

In 1st quarter 1982 dollars, product prices for the coal liquefaction complex are:

Propane	$42.00 per barrel or $264.20 per cubic metre
Butane	$42.00 per barrel or $264.20 per cubic metre
Stabilized Gasoline	$50.00 per barrel or $314.50 per cubic metre
Syncrude	$42.90 per barrel or $269.80 per cubic metre
Sulphur	$100.00 per tonne
Ammonia	$260.00 per tonne
Soda Ash	$65.00 per tonne

Annual product revenues are given in Sections 2.5 and 2.6 of this report.

Capital and Operating Cost Estimating

2.0 CAPITAL AND OPERATING COST ESTIMATES

2.1 CAPITAL COST SUMMARY - BASE CASE

Direct Costs	
Process Units	$777 000 000
Other Services	424 000 000
Total Direct Costs	$1 200 000 000
Indirect Costs	
Home Office, Field including construction camp, tools, equipment rental, taxes, fees (35% of direct cost)	$420 000 000
Contingency (25%)	400 000 000
Total Indirect Costs	$820 000 000
GRAND TOTAL	$2 020 000 000

BASE CASE

Direct Costs - Process Units

Area 200	Coal Receiving Preparation	$ 17 000 000
Area 300	Coal Liquefaction	385 700 000
Area 400	Atmospheric & Vacuum Distillation	31 200 000
Area 500	Naphtha Hydrotreater	12 000 000
Area 600	Middle Distillate Hydrotreater	30 900 000
Area 700	Heavy Distillate Hydrotreater	25 200 000
Area 800	Light Ends Recovery	31 200 000
Area 900	Hydrogen Production	177 300 000
Area 1000	Oxygen Plant	44 100 000
Area 1100	Sulphur Plant	22 400 000
	Total Direct Costs - Process Units	$777 000 000

BASE CASE

Direct Costs - Other Services

Area 1200 Steam & Power Generation	$ 79 700 000
Flue Gas Desulphurization	9 100 000
Area 1300 Phenol Recovery	3 100 000
Area 1400 Ammonia Recovery	13 100 000
Area 1500 Offsites	
- Catalyst & Chemical Storage	1 600 000
- Product Tankage & Load out	53 200 000
- Effluent Treatment	17 400 000
- Flare System	2 600 000
- Solid Waste Removal	9 000 000
Area 1600 Utilities	
- Plant & Instrument Air System	1 300 000
- Steam Distribution	4 000 000
- Cooling Tower System	6 000 000
- Boiler Feed Water System	5 100 000
- Fire Water System	400 000
- Raw Water Treatment	1 900 000
- Effluent & Fire Water Ponds	1 200 000
- Fuel Gas System	300 000
Miscellaneous Services	
- Interbattery Piping	49 300 000
- Site Development	117 700 000
- Buildings	19 700 000
- Site Access	2 000 000
- Catalyst, Chemicals & Royalties	25 300 000
Total Direct Costs - Other Services	$424 000 000

2.2 CAPITAL COST SUMMARY - ALTERNATIVE CASES

2.2.1 Case 1 - 50% Distillate Yield

Direct Costs

Process Units	$802 590 000
Other Services	419 570 000
Total Direct Costs	$1 222 160 000

Indirect Costs

Home Office, Field including construction camp, tools, equipment rental, taxes & fees (35% of direct cost)	$427 740 000
Contingency (25%)	$412 100 000
Total Indirect Costs	$839 840 000
GRAND TOTAL	$2 062 000 000

CASE 1

Direct Costs - Process Units

Area		
200	Coal Receiving & Preparation	$ 17 000 000
300	Coal Liquefaction	385 700 000
400	Atmospheric & Vacuum Distillation	35 700 000
500	Naphtha Hydrotreater	13 700 000
600	Middle Distillate Hydrotreater	35 300 000
700	Heavy Distillate Hydrotreater	28 800 000
800	Light Ends Recovery	31 200 000
900	Hydrogen Production	184 790 000
1000	Oxygen Plant	48 000 000
1100	Sulphur Plant	22 400 000
	Total Direct Costs - Process Units	$802 590 000

CASE 1

Direct Costs - Other Services

Area	
1200 Steam & Power Generation	$ 76 070 000
Flue Gas Desulphurization	Nil
1300 Phenol Recovery	3 100 000
1400 Ammonia Recovery	13 100 000
1500 Offsites	
- Catalyst & Chemical Storage	1 600 000
- Product Tankage & Load out	61 200 000
- Effluent Treatment	17 400 000
- Flare System	2 600 000
- Solid Waste Removal	9 000 000
1600 Utilities	
- Plant & Instrument Air System	1 300 000
- Steam Distribution	4 100 000
- Cooling Tower System	6 100 000
- Boiler Feed Water System	5 100 000
- Fire Water System	400 000
- Raw Water Treatment	1 900 000
- Effluent & Fire Water Ponds	1 200 000
- Fuel Gas System	300 000
Miscellaneous Services	
- Interbattery Piping	50 400 000
- Site Development	117 700 000
- Buildings	19 700 000
- Site Access	2 000 000
- Catalysts, Chemicals & Royalties	25 300 000
Total Direct Costs - Other Services	$ 419 570 000

2.2.2 Case 2 - Residue and Fuel Gas to MTG

Direct Costs	
Process Units	$1 093 700 000
Other Services	400 000 000
Total Direct Costs	$1 493 700 000
Indirect Costs	
Home Office, Field including construction camp, tools, equipment rental, taxes & fees (35% of direct cost)	$ 522 800 000
Contingency (25%)	503 500 000
Total Indirect Costs	$1 026 300 000
GRAND TOTAL	$2 520 000 000

CASE 2

Direct Costs - Process Units

Area		
200	Coal Receiving & Preparation	$ 17 000 000
300	Coal Liquefaction	385 700 000
400	Atmospheric & Vacuum Distillation	31 200 000
500	Naphtha Hydrotreater	12 000 000
600	Middle Distillate Hydrotreater	30 900 000
700	Heavy Distillate Hydrotreater	25 200 000
800	Light Ends Recovery	31 200 000
900	Hydrogen Production	265 600 000
1000	Oxygen Plant	62 700 000
1100	Sulphur Plant	23 200 000
1700	Fuel Gas Gasification	49 100 000
1800	Methanol Synthesis	86 400 000
1900	Methanol Conversion	57 400 000
2000	MTG Fractionation	13 300 000
2100	MTG Alkylation	2 800 000
	Total Direct Costs - Process Units	$1 093 700 000

CASE 2

Direct Costs - Other Services

Area	
1200 Steam & Power Generation	$ Nil
Flue Gas Desulphurization	Nil
1300 Phenol Recovery	3 100 000
1400 Ammonia Recovery	13 100 000
1500 Offsites	
- Catalyst & Chemical Storage	1 600 000
- Product Tankage & Load out	60 800 000
- Effluent Treatment	17 400 000
- Flare System	2 600 000
- Solid Waste Removal	5 200 000
1600 Utilities	
- Plant & Instrument Air System	1 500 000
- Steam Distribution	4 600 000
- Cooling Tower System	7 900 000
- Boiler Feed Water System	5 400 000
- Fire Water System	600 000
- Raw Water Treatment	2 400 000
- Effluent & Fire Water Ponds	1 200 000
- Fuel Gas System	300 000
Miscellaneous Services	
- Interbattery Piping	60 700 000
- Site Development	144 300 000
- Buildings	24 000 000
- Site Access	2 000 000
- Catalysts, Chemicals & Royalties	32 500 000
- Power Transmission	8 800 000
Total Direct Costs - Other Services	$400 000 000

2.2.3 Case 3 - Coal Tailings, Residue & Fuel Gas to MTG

Direct Cost

Process Units	$1 134 405 000
Other Services	425 000 000
Total Direct Costs	$1 559 405 000

Indirect Cost

Home Office, Field including construction camp, tools, equipment rental, taxes & fees (35% of direct cost)	$ 545 100 000
Contingency (25%)	$ 525 495 000
Total Indirect Cost	$1 070 595 000
GRAND TOTAL	$2 630 000 000

CASE 3

Direct Costs - Process Units

Area		
200	Coal Receiving & Preparation	$ 19 000 000
300	Coal Liquefaction	385 700 000
400	Atmospheric & Vacuum Distillation	31 200 000
500	Naphtha Hydrotreater	12 000 000
600	Middle Distillate Hydrotreater	30 900 000
700	Heavy Distillate Hydrotreater	25 200 000
800	Light Ends Recovery	31 200 000
900	Hydrogen Production	275 345 000
1000	Oxygen Plant	67 700 000
1100	Sulphur Plant	23 200 000
1700	Fuel Gas Gasification	49 100 000
1800	Methanol Synthesis	99 360 000
1900	Methanol Conversion	66 000 000
2000	MTG Fractionation	15 300 000
2100	MTG Alkylation	3 200 000
	Total Direct Costs - Process Units	$1 134 405 000

CASE 3

Direct Costs - Other Services

Area	
1200 Steam & Power Generation	$ Nil
Flue Gas Desulphurization	Nil
1300 Phenol Recovery	3 100 000
1400 Ammonia Recovery	13 100 000
1500 Offsites	
- Catalyst & Chemical Storage	1 600 000
- Product Tankage & Load out	69 160 000
- Effluent Treatment	17 400 000
- Flare System	2 600 000
- Solid Waste Removal	12 600 000
1600 Utilities	
- Plant & Instrument Air System	1 500 000
- Steam Distribution	5 100 000
- Cooling Tower System	8 690 000
- Boiler Feed Water System	5 800 000
- Fire Water System	800 000
- Raw Water Treatment	3 000 000
- Effluent & Fire Water Ponds	1 200 000
- Fuel Gas System	300 000
Miscellaneous Services	
- Interbattery Piping	63 450 000
- Site Development	144 300 000
- Buildings	24 000 000
- Site Access	2 000 000
- Catalysts, Chemicals & Royalties	36 500 000
- Power Transmission	8 800 000
Total Direct Costs - Other Services	$425 000 000

2.2.4 Case 4 - Steam- Methane Reforming, Coal Tailings & Residue to MTG

Direct Cost

Process Units	$ 1 360 250 000
Other Services	440 560 000
Total Direct Costs	$ 1 800 810 000

Indirect Costs

Home Office, Field including construction camp, tools, equipment rental, taxes & fees (35% of direct cost)	$ 630 190 000
Contingency (25%)	608 000 000
Total Indirect Cost	$ 1 238 190 000
GRAND TOTAL	$ 3 039 000 000

CASE 4

Direct Costs - Process Units

Area		
200	Coal Receiving & Preparation	$ 19 000 000
300	Coal Liquefaction	385 700 000
400	Atmospheric & Vacuum Distillation	31 200 000
500	Naphtha Hydrotreater	12 000 000
600	Middle Distillate Hydrotreater	30 900 000
700	Heavy Distillate Hydrotreater	25 200 000
800	Light Ends Recovery	31 200 000
900	Steam-Methane Hydrogen Production	235 600 000
1000	Oxygen Plant	60 630 000
1100	Sulphur Plant	23 200 000
1700	Residue Gasification	270 770 000
1800	Methanol Synthesis	129 600 000
1900	Methanol Conversion	81 100 000
2000	MTG Fractionation	19 950 000
2100	MTG Alkylation	4 200 000
	Total Direct Costs - Process Units	$ 1 360 250 000

CASE 4

Direct Costs - Other Services

Area	
1200 Steam & Power Generation	$ Nil
Flue Gas Desulphurization	Nil
1300 Phenol Recovery	3 100 000
1400 Ammonia Recovery	13 100 000
1500 Offsites	
- Catalyst & Chemical Storage	1 600 000
- Product Tankage & Load Out	70 900 000
- Effluent Treatment	20 000 000
- Flare System	2 600 000
- Solid Waste Removal	13 600 000
1600 Utilities	
- Plant & Instrument Air System	1 500 000
- Steam Distribution	5 000 000
- Cooling Tower System	11 610 000
- Boiler Feed Water System	6 000 000
- Fire Water System	800 000
- Raw Water Treatment	2 200 000
- Effluent & Fire Water Ponds	1 200 000
- Fuel Gas System	300 000
Miscellaneous Services	
- Interbattery Piping	68 450 000
- Site Development	144 300 000
- Buildings	24 000 000
- Site Access	2 000 000
- Catalyst, Chemicals & Royalties	39 500 000
- Power Transmission	8 800 000
Total Direct Cost - Other Services	$440 560 000

2.2.5 Case 5 - Steam- Methane Reforming

Direct Cost	
Process Units	$ 778 500 000
Other Services	448 000 000
Total Direct Costs	$ 1 236 500 000
Indirect Costs	
Home Office, Field including construction camp, tools, equipment rental, taxes & fees (35% of direct cost)	$ 433 000 000
Contingency (25%)	417 500 000
Total Indirect Cost	$ 850 500 000
GRAND TOTAL	$ 2 087 000 000

CASE 5

Direct Costs - Process Units

Area		
200	Coal Receiving & Preparation	$ 17 000 000
300	Coal Liquefaction	385 700 000
400	Atmospheric & Vacuum Distillation	31 200 000
500	Naphtha Hydrotreater	12 000 000
600	Middle Distillate Hydrotreater	30 900 000
700	Heavy Distillate Hydrotreater	25 200 000
800	Light Ends Recovery	31 200 000
900	Steam-Methane Hydrogen Production	235 600 000
1100	Sulphur Plant	19 700 000
	Total Direct Costs - Process Units	$ 788 500 000

CASE 5

Direct Costs - Other Services

Area	
1200 Steam & Power Generation	$100 660 000
Flue Gas Desulphurization	13 500 000
1300 Phenol Recovery	3 100 000
1400 Ammonia Recovery	13 100 000
1500 Offsites	
- Catalyst & Chemical Storage	1 600 000
- Product Tankage & Load Out	53 200 000
- Effluent Treatment	17 400 000
- Flare System	2 600 000
- Solid Waste Removal	9 000 000
1600 Utilities	
- Plant & Instrument Air System	1 300 000
- Steam Distribution	4 100 000
- Cooling Tower System	4 800 000
- Boiler Feed Water System	5 800 000
- Fire Water System	400 000
- Raw Water Treatment	1 900 000
- Effluent & Fire Water Ponds	1 200 000
- Fuel Gas System	300 000
Miscellaneous Services	
- Interbattery Piping	49 300 000
- Site Development	117 700 000
- Buildings	19 700 000
- Site Access	2 000 000
- Catalyst, Chemicals & Royalties	25 340 000
Total Direct Cost - Other Services	$448 000 000

2.3 ANNUAL OPERATING COSTS - BASE CASE

2.3.1 Manpower

At full capacity the coal liquefaction complex employs a staff of 860 personnel at a cost of $32 680 000 per year.

2.3.2 Raw Materials

At full production the complex consumes:

- 10 000 tonne/day of coal @ $80.00 per tonne

Annual consumption is based on 310 operating days per year.

2.3.3 Products

At full production the plant produces:

- 3726 barrels/day of propane	@ $42.00 per barrel
- 2942 barrels/day of butane	@ $42.00 per barrel
- 24280 barrels/day of syncrude	@ $42.90 per barrel

consisting of:

9045 barrel/day naphtha
11020 barrel/day middle distillate
4215 barrel/day heavy distillate

- 35 tonne/day of sulphur	@ $100.00 per tonne
- 115 tonne/day of anhydrous ammonia	@ $260.00 per tonne
- 9 tonne/day of soda ash	@ $ 65.00 per tonne
- 423200 Nm³/day of excess fuel gas	@ $0.131/Nm³ (After 12% discount for heating value)

Annual production is based on 310 operating days value per year.

2.3.4 Catalysts and Chemicals

At full production the plant utilizes catalysts and chemicals as summarized below:

Coal Conversion Catalyst	$ 40 212 000
Diglycol Amine	11 000
Naphtha Catalyst	283 000
Middle Distillate Catalyst	277 000
Heavy Distillate Catalyst	143 000
PSA & Molecular Sieve Absorbants	74 000
Sour Shift Catalyst	567 000
Acid Gas Solvent	203 000
Sulphur Plant Chemicals	139 000
Lime (Flue Gas Scrubbing)	366 000
Phosphoric Acid	4 000
Cooling Water Chemicals	218 000
Potable Water Chemicals	21 000
Caustic	358 000
Sulphuric Acid	47 000
Activated Carbon	11 000
Resins for BFW Plant	45 000
	$ 42 979 000
Allowance for minor items	521 000
	$ 43 500 000

Annual consumption is based on 310 days per year.

2.3.5 Maintenance and Modifications

At full production, the maintenance budget for part is about one percent of the fixed capital investment. Maintenance labour is included in manpower requirements.

2.3.6 Solid Waste Removal

At full production the base case complex will produce about 1680 tonne per day of solid wastes comprising excess residue, slag, ash and solid waste from flue gas desulphurization. The cost of disposal of this waste offsite is estimated to be $10.00 per tonne.

ANNUAL OPERATING COST SUMMARY

Base Case	**1982 Dollars**
Raw Materials - Coal	$248 000 000
Catalysts and Chemicals	43 500 000
Labour and Overhead	32 680 000
Maintenance Parts (1% of FCI)	20 200 000
Maintenance*	2 000 000
Insurance and Taxes	41 500 000
Solid Waste Removal	5 200 000
	$393 080 000

*Subcontracted Maintenance.

2.4 ANNUAL OPERATING COST SUMMARY - ALTERNATIVE CASES

2.4.1 Case 1 - 50% Distillate Yields

	1982 dollars
Raw Materials - Coal	$248 000 000
Catalysts and Chemicals	43 800 000
Labour and Overhead (860 personnel)	32 680 000
Maintenance Parts	20 950 000
Maintenance*	2 095 000
Insurance & Taxes	42 500 000
Solid Waste Removal	1 700 000
Annual Operating Cost	$391 725 000

*Subcontracted Maintenance.

2.4.2 Case 2 - Residue & Fuel Gas to MTG

	1982 dollars
Raw Material - Coal	$248 000 000
Catalysts and Chemicals	46 500 000
Labour and Overhead (990 personnel)	37 620 000
Maintenance Parts	25 200 000
Maintenance*	2 000 000
Electricity 1 763 280 Mwh/yr @ $22.60/Mwh	39 850 000
Insurance & Taxes	51 800 000
Solid Waste Removal	1 700 000
Annual Operating Cost	$452 670 000

*Subcontracted Maintenance.

2.4.3 Case 3 - Coal Tailings, Residue and Fuel Gas to MTG

	1982 dollars
Raw Materials - Coal	$248 000 000
- Coal Tailings	Nil
Catalysts and Chemicals	48 500 000
Labour and Overhead (1040 personnel)	39 520 000
Maintenance Parts	26 300 000
Maintenance*	2 000 000
Electricity 2 284 080 Mwh/yr @ $22.60/Mwh	51 620 000
Insurance & Taxes	54 000 000
Solid Waste Removal	2 500 000
Annual Operating Cost	$472 440 000

*Subcontracted Maintenance.

2.4.4 Case 4 - Steam-Methane Reforming, Coal Tailings & Residue to MTG

	1982 dollars
Raw Materials - Coal	\$248 000 000
- Coal Tailings	Nil
- Natural Gas 815.1 x 10^6 Nm³/year	121 400 000
Catalyst and Chemicals	50 500 000
Labour and Overhead (1040 personnel)	39 520 000
Maintenance Parts	30 970 000
Maintenance*	2 000 000
Electricity 1 159 150 Mwh/yr @ \$22.60/Mwh	26 200 000
Insurance & Taxes	63 550 000
Solid Waste Removal	2 500 000
Annual Operating Cost	\$584 640 000

*Subcontracted Maintenance.

2.4.5 Case 5 - Steam-Methane Reforming

	1982 dollars
Raw Materials - Coal	$248 000 000
- Natural Gas 699.2 x 10^6 Nm³/year	104 200 000
Catalysts and Chemicals	43 500 000
Labour and Overhead (860 personnel)	32 680 000
Maintenance Parts	20 870 000
Maintenance*	2 000 000
Insurance & Taxes	42 800 000
Solid Waste Removal	Nil
Annual Operating Cost	$494 050 000

*Subcontracted Maintenance.

2.5 ANNUAL REVENUE - BASE CASE

	1982 Dollars
Propane	$ 48 510 000
Butane	38 305 000
Syncrude	322 900 000
Sulphur	1 075 000
Ammonia	9 350 000
Soda Ash	180 000
Fuel Gas Credit	17 180 000
Total Revenue (1982 dollars)	$437 500 000
less Annual Operating Costs	393 080 000
Gross Margin	$ 44 420 000

WORKING CAPITAL - BASE CASE

Coal Storage (15 days)	$ 12 000 000
Syncrude Storage (30 days)	31 250 000
Propane Storage (30 days)	4 690 000
Butane Storage (30 days)	3 700 000
Sulphur Storage (60 days)	200 000
Ammonia Storage (30 days)	900 000
Soda Ash Storage (90 days)	50 000
Labour Payroll	2 700 000
	$ 55 490 000

2.6 ANNUAL REVENUE - ALTERNATIVE CASES

2.6.1 Case 1 - 50% Distillate Yield

	1982 Dollars
Propane (3725 barrels/day)	\$ 48 500 000
Butane (2940 barrels/day)	38 280 000
Syncrude (28665 barrels/day)	381 120 000
Sulphur (36 tonne/day)	1 150 000
Ammonia (116 tonne/day)	9 350 000
Soda Ash (9 tonne/day)	180 000
Fuel Gas Credit (380 880 Nm³/day)	15 470 000
Total Revenue	\$494 050 000
Less Annual Operating Costs	391 725 000
Gross Margin	\$102 325 000

Working Capital - Case 1

Coal Storage (15 days)	\$ 12 000 000
Syncrude Storage (30 days)	36 900 000
Propane Storage (30 days)	4 690 000
Butane Storage (30 days)	3 700 000
Sulphur Storage (60 days)	200 000
Ammonia Storage (30 days)	900 000
Soda Ash Storage (90 days)	50 000
Labour Payroll	2 700 000
	\$ 61 140 000

2.6.2 Case 2 - Residue and Fuel Gas to MTG

	1982 Dollars
Propane (4190 barrels/day)	$ 54 555 000
Butane (3640 barrels/day)	47 395 000
MTG Gasoline (8210 barrels/day)	127 255 000
Syncrude (24280 barrels/day)	322 900 000
Sulphur (35 tonne/day)	1 075 000
Ammonia (116 tonne/day)	9 350 000
Soda Ash (9 tonne/day)	180 000
Total Revenue	$562 710 000
less Annual Operating Costs	452 670 000
Gross Margin	$110 040 000

Working Capital - Case 2

Coal Storage (15 days)	$ 12 000 000
Syncrude Storage (30 days)	31 250 000
MTG Gasoline Storage (30 days)	12 315 000
Butane Storage (30 days)	4 590 000
Propane Storage (30 days)	5 280 000
Sulphur Storage (60 days)	200 000
Ammonia Storage (30 days)	900 000
Soda Ash Storage (90 days)	50 000
Labour Payroll	3 100 000
	$ 69 685 000

2.6.3 Case 3 - Tailings, Residue and Fuel Gas to MTG

	1982 Dollars
Propane (4350 barrels/day)	$ 56 640 000
Butane (3900 barrels/day)	50 780 000
MTG Gasoline (10210 barrels/day)	158 255 000
Syncrude (24280 barrels/day)	322 900 000
Sulphur (35 tonne/day)	1 075 000
Ammonia (116 tonne/day)	9 350 000
Soda Ash (9 tonne/day)	180 000
Total Revenue	$599 180 000
Less Annual Operating Costs	472 440 000
Gross Margin	126 740 000

Working Capital

Coal Storage (15 days)	$ 12 000 000
Syncrude Storage (30 days)	31 250 000
MTG Gasoline Storage (30 days)	15 315 000
Propane Storage (30 days)	5 480 000
Butane Storage (30 days)	4 915 000
Sulphur Storage (60 days)	200 000
Ammonia Storage (30 days)	900 000
Soda Ash Storage (90 days)	50 000
Labour Payroll	3 250 000
	$ 73 360 000

2.6.4 Case 4 - Steam-Methane Reforming, Coal Tailings & Residue to MTG

	1982 Dollars
Propane (4665 barrels/day)	$ 22 225 000
Butane (4420 barrels/day)	57 550 000
MTG Gasoline (17 070 barrels/day)	264 590 000
Syncrude (24 280 barrels/day)	322 900 000
Sulphur (35 tonne/day)	1 075 000
Ammonia (116 tonne/day)	9 350 000
Soda Ash (9 tonne/day)	180 000
Total Revenue	$677 870 000
Less Annual Operating Costs	584 640 000
Gross Margin	$ 93 230 000

Working Capital

Coal Storage (15 days)	$ 12 000 000
Syncrude Storage (30 days)	31 200 000
MTG Gasoline Storage (30 days)	25 600 000
Propane Storage (30 days)	2 150 000
Butane Storage (30 days)	5 550 000
Sulphur Storage (60 days)	200 000
Ammonia Storage (30 days)	900 000
Soda Ash Storage (90 days)	50 000
Labour Payroll	3 250 000
	$ 80 900 000

2.6.5 Case 5 - Steam-Methane Reforming

	1982 Dollars
Propane (3726 barrels/day)	$ 48 510 000
Butane (2942 barrels/day)	38 305 000
Syncrude (24 280 barrels/day)	322 900 000
Sulphur (28 tonne/day)	868 000
Ammonia (116 tonne/day)	9 350 000
Soda Ash (9 tonne/day)	180 000
Total Revenue	$420 113 000
Less Annual Operating Costs	494 050 000
Gross Margin	[73 937 000] [negative]

Working Capital

Coal Storage (15 days)	$ 12 000 000
Syncrude Storage (30 days)	31 250 000
Propane Storage (30 days)	4 690 000
Butane Storage (30 days)	3 700 000
Sulphur Storage (60 days)	160 000
Ammonia Storage (30 days)	900 000
Soda Ash Storage (90 days)	50 000
Labour Payroll	2 700 000
	$ 55 450 000

2.7 PROCESS EQUIPMENT ESTIMATE SUMMARY

BASE CASE

PROCESS EQUIPMENT

ESTIMATE SUMMARY

CLIENT: Western Canada Coal Liquefaction Study DATE: February 1982

TYPE OF ESTIMATE: Order of Magnitude

AREA: 200 - Coal Receiving & Handling

PROCESS PACKAGE EQUIPMENT

TOTAL DIRECT COST

Conveyors, Stackers, Reclaimers and associated equipment	$ 15 000 000
Feed Surge Bins and associated equipment	2 000 000

TOTAL DIRECT COST PROCESS EQUIPMENT	$ 17 000 000

Cost Basis

The estimate is based on in-house Kilborn information.

PROCESS EQUIPMENT

ESTIMATE SUMMARY

CLIENT: Western Canada Coal Liquefaction Study DATE: February 1982

TYPE OF ESTIMATE: Order of Magnitude

AREA: 300 - Coal Liquefaction

EQUIPMENT	INSTALLED COST	
Pumps	18 990 000	
Compressors	11 290 000	
Tanks		
Drums	42 306 000	
Reactors	58 800 000	
Towers		
Exchangers	19 675 000	
Fired Heaters	23 076 000	
Mechanical Equipment		
Package Systems	9 537 000	
Other		
TOTAL PROCESS EQUIPMENT INSTALLED COST (PEIC)	183 674 000	
Piping 60 % of PEIC	110 200 000	
Instrumentation 20 % of PEIC	36 726 000	
Electrical 15 % of PEIC	27 550 000	
Bldg./Structural 15 % of PEIC	27 550 000	
SUBTOTAL		$102 026 000
TOTAL DIRECT COST PROCESS EQUIPMENT		$ 385 700 000

Direct Cost Basis

The estimate was factored from estimated major equipment prices in either Canadian or U.S. Gulf Coast dollars. Factors were applied to convert U.S. Gulf Coast to Canadian dollars Greenhills location.

U.S. $ Gulf Coast x Conversion Factor = Canadian Dollars Greenhills location
Conversion Factor = 0.75 (domestic equipment) + 0.58 (foreign equipment).

PROCESS EQUIPMENT

ESTIMATE SUMMARY

CLIENT: Western Canada Coal Liquefaction Study DATE: February 1982

TYPE OF ESTIMATE: Order of Magnitude

AREA: 400 - Atmospheric & Vacuum Distillation

EQUIPMENT	INSTALLED COST	
Pumps	881 000	
Compressors		
Tanks		
Drums	178 000	
Reactors		
Towers	3 040 000	
Exchangers	2 897 000	
Fired Heaters	6 063 000	
Mechanical Equipment		
Package Systems	1 806 000	
Other		
TOTAL PROCESS EQUIPMENT INSTALLED COST (PEIC)	14 865 000	
Piping 60 % of PEIC	890.500	
Instrumentation 20 % of PEIC	2 970 000	
Electrical 15 % of PEIC	2 230 000	
Bldg./Structural 15 % of PEIC	2 230 000	
SUBTOTAL		$ 16 335 000
TOTAL DIRECT COST PROCESS EQUIPMENT		$ 31 200 000

Direct Cost Basis

The estimate was factored from estimated major equipment prices in either Canadian or U.S. Gulf Coast dollars. Factors were applied to convert U.S. Gulf Coast to Canadian dollars Greenhills location.

U.S. $ Gulf Coast x Conversion Factor = Canadian Dollars Greenhills location
Conversion Factor = 0.75 (domestic equipment) + 0.58 (foreign equipment).

PROCESS EQUIPMENT

ESTIMATE SUMMARY

CLIENT: Western Canada Coal Liquefaction Study DATE: February 1982

TYPE OF ESTIMATE: Order of Magnitude

AREA: 500 - Naphtha Hydrotreater

PROCESS PACKAGE EQUIPMENT

TOTAL DIRECT COST

Naphtha Hydrotreater $ 8 353 000 US x 1.44 = $12 000 000

TOTAL DIRECT COST PROCESS EQUIPMENT $ 12 000 000

Direct Cost Basis

The estimate is based on information from UOP Process Division of Des Plaines, Illinois.

U.S. $ Gulf Coast x Conversion Factor = Canadian $ Greenhills location

Conversion Factor = 0.94 material component + 0.50 labour component

= 1.44

PROCESS EQUIPMENT

ESTIMATE SUMMARY

CLIENT: Western Canada Coal Liquefaction Study DATE: February 1982

TYPE OF ESTIMATE: Order of Magnitude

AREA: 600 - Middle Distillate Hydrotreater

PROCESS PACKAGE EQUIPMENT

TOTAL DIRECT COST

Middle Distillate Hydrotreater	$ 21 460 000 US x 1.44 = $ 30 900 000

TOTAL DIRECT COST PROCESS EQUIPMENT	$ 30 900 000

Direct Cost Basis

The estimate is based on information from UOP Process Division of Des Plaines, Illinois.

U.S. $ Gulf Coast x Conversion Factor = Canadian $ Greenhills location

Conversion Factor = 0.94 material component + 0.50 labour component
= 1.44

PROCESS EQUIPMENT

ESTIMATE SUMMARY

CLIENT: Western Canada Coal Liquefaction Study DATE: February 1982

TYPE OF ESTIMATE: Order of Magnitude

AREA: 700 - Heavy Distillate Hydrotreater

PROCESS PACKAGE EQUIPMENT

TOTAL DIRECT COST

Heavy Distillate Hydrotreater	$ 17 495 000 US x 1.44 = $ 25 200 000

TOTAL DIRECT COST PROCESS EQUIPMENT	$25 200 000

Direct Cost Basis

The estimate is based on information from UOP Process Division of Des Plaines, Illinois.

U.S. $ Gulf Coast x Conversion Factor = Canadian $ Greenhills location

Conversion Factor = 0.94 material component + 0.50 labour component
= 1.44

PROCESS EQUIPMENT

ESTIMATE SUMMARY

CLIENT: Western Canada Coal Liquefaction Study DATE: February 1982

TYPE OF ESTIMATE: Order of Magnitude

AREA: 800 - Light Ends Recovery

PROCESS PACKAGE EQUIPMENT

TOTAL DIRECT COST

CO_2 Removal	$ 2 830 000 US x 1.44=	$ 4 070 000
Light Ends Separation	$12 632 000 US x 1.44=	$ 18 170 000
Cryogenic & PSA Recovery Unit	$ 6 225 000 US x 1.44=	$ 8 960 000

TOTAL DIRECT COST PROCESS EQUIPMENT $ 31 200 000

Direct Cost Basis

This estimate is based on Kilborn in-house information and Union Carbide, Linde Division information.

U.S. $ Gulf Coast x Conversion Factor = Canadian $ Greenhills location

Conversion Factor = 0.94 material component + 0.50 labour component
= 1.44

PROCESS EQUIPMENT

ESTIMATE SUMMARY

CLIENT: Western Canada Coal Liquefaction Study DATE: February 1982

TYPE OF ESTIMATE: Order of Magnitude

AREA: 900 - Sulphur Plant

PROCESS PACKAGE EQUIPMENT

TOTAL DIRECT COST

Stretford Sulphur Plant $13 560 000 US x 1.44 = $ 19 526 000

AUXILLARY EQUIPMENT

INSTALLED COST --	Pumps	11 000	
	Compressors		
	Tanks	11 000	
	Drums		
	Exchangers		
	Mechanical Equip.		
	Package Systems	1 340 000	
	Other		
TOTAL AUX. EQUIP. INSTALLED COST		1 362 000	
PIPING, INST.& ELEC. @ 110%		1 512 000	
SUBTOTAL			$ 2 874 000
TOTAL DIRECT COST PROCESS EQUIPMENT			$ 22 400 000

Direct Cost Basis

This estimate is based on in-house Kilborn information and information from Peabody Process Systems of Stamford, Connecticut.

U.S. $ Gulf Coast x Conversion Factor = Canadian $ Greenhills location

Conversion Factor = 0.94 material component + 0.50 labour component
= 1.44

PROCESS EQUIPMENT

ESTIMATE SUMMARY

CLIENT: Western Canada Coal Liquefaction Study DATE: February 1982

TYPE OF ESTIMATE: Order of Magnitude

AREA: 1000 - Hydrogen Production

PROCESS PACKAGE EQUIPMENT

TOTAL DIRECT COST

Gasification & Quench	$64 275 000 US x 1.44=	$ 92 590 000
Water-Gas Shift	$12 044 000 US x 1.44=	17 350 000
Acid Gas Removal	$46 776 000 US x 1.44=	67 360 000

TOTAL DIRECT COST PROCESS EQUIPMENT $ 177 300 000

Direct Cost Basis

This estimate is based on Kilborn in-house information.

U.S. $ Gulf Coast x Conversion Factor = Canadian $ Greenhills location

Conversion Factor = 0.94 material component + 0.50 labour component
= 1.44

PROCESS EQUIPMENT

ESTIMATE SUMMARY

CLIENT: Western Canada Coal Liquefaction Study DATE: February 1982

TYPE OF ESTIMATE: Order of Magnitude

AREA: 1100 - Oxygen Plant

PROCESS PACKAGE EQUIPMENT

TOTAL DIRECT COST

Oxygen Separation Plant	$30 596 000 US x 1.44 =	$ 44 100 000

TOTAL DIRECT COST PROCESS EQUIPMENT	$ 44 100 000

Direct Cost Basis

This estimate is based on Kilborn in-house information.

U.S. $ Gulf Coast x Conversion Factor = Canadian $ Greenhills location

Conversion Factor = 0.94 material component + 0.50 labour component
= 1.44

PROCESS EQUIPMENT

ESTIMATE SUMMARY

CLIENT: Western Canada Coal Liquefaction Study DATE: February 1982

TYPE OF ESTIMATE: Order of Magnitude

AREA: 1200 - Steam & Power Generation

PROCESS PACKAGE EQUIPMENT

TOTAL DIRECT COST

Steam Boilers	$ 17 801 000 US x 1.9 =		$ 33 840 000
Ash Blending	1 909 000 US x 1.9 =		3 630 000
Power Generation			
Turbines		$ 11 017 000	
Exchangers		8 600 000	
Transformers & Breakers		481 000	
Process Equip. Cost (PEC)		20 058 000	
Installation (PEC x 1.1)		22 132 000	42 230 000

TOTAL DIRECT COST PROCESS EQUIPMENT $ 79 700 000

Direct Cost Basis

This estimate is based on information from Babcock & Wilcox of Cambridge, Ont. and Canadian General Electric of Toronto, Ontario

PROCESS EQUIPMENT

ESTIMATE SUMMARY

CLIENT: Western Canada Coal Liquefaction Study DATE: February 1982

TYPE OF ESTIMATE: Order of Magnitude

AREA: 1200 - Flue Gas Desulphurization (Steam & Power Generation)

PROCESS PACKAGE EQUIPMENT

TOTAL DIRECT COST $ 5 778 000 US x 1.44 = $ 8 300 000

AUXILLARY EQUIPMENT

INSTALLED COST --	Pumps	4 000	
	Compressors	274 000	
	Tanks		
	Drums		
	Exchangers		
	Mechanical Equip.	96 000	
	Package Systems		
	Other		
TOTAL AUX. EQUIP. INSTALLED COST		374 000	
PIPING,INST.& ELEC. @ 110%		426 000	800 000

SUBTOTAL

TOTAL DIRECT COST PROCESS EQUIPMENT $ 9 100 000

Direct Cost Basis

This estimate is based on Kilborn in-house information and information from Wheelebrator-Frye of Pittsburgh, Pennsylvania.

U.S. $ Gulf Coast x Conversion Factor = Canadian $ Greenhills location

Conversion Factor = 0.94 material component + 0.50 labour component
= 1.44

PROCESS EQUIPMENT

ESTIMATE SUMMARY

CLIENT: Western Canada Coal Liquefaction Study DATE: February 1982

TYPE OF ESTIMATE: Order of Magnitude

AREA: 1300 - Phenol Recovery

PROCESS PACKAGE EQUIPMENT

TOTAL DIRECT COST $ 2 143 000 US x 1.44 = $ 3 075 000

AUXILLARY EQUIPMENT

INSTALLED COST -	Pumps		
	Compressors	4 000	
	Tanks		
	Drums		
	Exchangers	8 000	
	Mechanical Equip.		
	Package Systems		
	Other		
TOTAL AUX. EQUIP. INSTALLED COST		12 000	
PIPING,INST.& ELEC. @ 110%		13 000	$ 25 000

SUBTOTAL

TOTAL DIRECT COST PROCESS EQUIPMENT $ 3 100 000

Direct Cost Basis

This estimate is based on Kilborn in-house information and information from Koppers Co. of Pittsburgh, Pennsylvania.

U.S. $ Gulf Coast x Conversion Factor = Canadian $ Greenhills location

Conversion Factor = 0.94 material component + 0.50 labour component
= 1.44

PROCESS EQUIPMENT

ESTIMATE SUMMARY

CLIENT: Western Canada Coal Liquefaction Study DATE: February 1982

TYPE OF ESTIMATE: Order of Magnitude

AREA: 1400 - Ammonia Recovery

PROCESS PACKAGE EQUIPMENT

TOTAL DIRECT COST $ 8 229 000 US x 1.44 = $ 11 850 000

AUXILLARY EQUIPMENT

INSTALLED COST -	Pumps		
	Compressors		
	Tanks	580 000	
	Drums		
	Exchangers		
	Mechanical Equip.		
	Package Systems		
	Other		
TOTAL AUX. EQUIP. INSTALLED COST		580 000	
PIPING,INST.& ELEC. @ 110%		670 000	$ 1 250 000

SUBTOTAL

TOTAL DIRECT COST PROCESS EQUIPMENT $ 13 100 000

Direct Cost Basis

This estimate is based on Kilborn in-house information and information from Chevron Research Co. of San Francisco, California.

U.S. $ Gulf Coast x Conversion Factor = Canadian $ Greenhills location

Conversion Factor = 0.94 material component + 0.50 labour component
= 1.44

PROCESS EQUIPMENT

ESTIMATE SUMMARY

CLIENT: Western Canada Coal Liquefaction Study DATE: February 1982

TYPE OF ESTIMATE: Order of Magnitude

AREA: 1500 - Offsites

PROCESS PACKAGE EQUIPMENT

TOTAL DIRECT COST

Catalyst & Chemical Storage	$ 1 600 000
Product Tankage (& load out)	53 200 000
Effluent Treatment	17 400 000
Flare System	2 600 000
Solid Waste Removal	9 000 000

TOTAL DIRECT COST PROCESS EQUIPMENT $ 82 500 000

Direct Cost Basis

This estimate is based on Kilborn in-house information.

PROCESS EQUIPMENT

ESTIMATE SUMMARY

CLIENT: Western Canada Coal Liquefaction Study DATE: February 1982

TYPE OF ESTIMATE: Order of Magnitude

AREA: 1600 - Utilities

PROCESS PACKAGE EQUIPMENT

TOTAL DIRECT COST

Plant & Instrument Air System	$ 1 300 000
Steam Distribution System	4 100 000
Cooling Tower System	6 100 000
Boiler Feed Water System	5 100 000
Fire Water System	400 000
Raw Water Treatment	1 900 000
Effluent & Fire Water Ponds	1 200 000
Fuel Gas System	300 000

TOTAL DIRECT COST PROCESS EQUIPMENT $ 20 400 000

Direct Cost Basis

This estimate is based on Kilborn in-house information.

2.8 EQUIPMENT PRICE SUMMARY

BASE CASE

KILBORN **EQUIPMENT** 2-52

CLIENT. Western Canada Coal Liquefaction Study

EQUIPMENT TYPE. Compressors

PROJ. Nº. 3349 DIV. NO. 15

AREA 300 - Coal Liquefaction

SHEET Nº ____ CONTº ON ____

REV. Nº

DATE REVISED: ____

ISSUE DATE February 1982

EQUIPMENT Nº		DESCRIPTION	ESTIMATED PRICE
Item No.	Quantity		
	2	Intermediate H_2 recycle compressor	$ 8 772 000
	2	H_2 recycle compressor	1 492 000
		Equipment Cost	10 264 000
		Installation at 10%	1 026 000
		Total Installed Cost	$ 11 290 000

KILBORN EQUIPMENT

2-53

CLIENT. Western Canada Coal Liquefaction Study

EQUIPMENT TYPE. Heat Exchangers

PROJ. Nº 3349 DIV. NO. 15

AREA 300 - Coal Liquefaction

SHEET Nº ____ CONTº ON ____

REV Nº

DATE REVISED: ____

ISSUE DATE February 1982

EQUIPMENT Nº		DESCRIPTION	ESTIMATED PRICE
Item No.	Quantity		
	5	Vapour Cooler/Fractionator Feed Heater	$ 385 000
	5	Vapour Final Cooler	555 000
	5	Vapour Cooler/Recycle Gas Heater	5 100 000
	5	Vapour Cooler/Fractionator Feed Heater	2 125 000
	5	Feed Cooler/Waste Heat Boiler	1 440 000
	5	Feed Cooler/Waste Heat Boiler	1 900 000
	5	Vapour Cooler/Waste Heat Boiler	1 825 000
	5	Vapour Condenser	333 000
	5	Vapour Cooler	279 000
	5	Quench Cooler	1 144 000
	5	Feed Final Cooler	2 201 000
	5	Vapour Cooler	931 000
		Equipment Cost	$ 18 218 000
		Installation at 8%	1 457 000
		Total Installed Cost	$ 19 675 000

KILBORN EQUIPMENT

2-54

CLIENT. Western Canada Coal Liquefaction Study

EQUIPMENT TYPE. Pumps

PROJ. Nº. 3349 DIV. NO. 15

AREA 300 - Coal Liquefaction

SHEET Nº ____ CONT'D ON ____

REV Nº.

DATE REVISED: ____

ISSUE DATE February 1982

EQUIPMENT Nº		DESCRIPTION	ESTIMATED PRICE
Item No.	Quantity		
	10	Slurry Circulation	585 000
	2	Condensate Booster	4 000
	2	Slurry Oil Booster	109 000
	10	Slurry Circulation	2 400 000
	15	Slurry Feed	5 445 000
	5	Ebullating Pump	3 797 000
	2	Water Circulation	308 000
	10	Slurry Circulation	293 000
	10	Fractionator Pump	208 000
	10	Fractionator Feed	507 000
	10	Quench Pump	66 000
	15	Slurry Oil Recycle	589 000
	15	Hydroclone Overflow Circulation	392 000
	10	Hydroclone Feed	174 000
	10	Hydroclone Underflow Circulation	162 000
	10	Sour Water Discharge	144 000
	10	Wet Oil Discharge	41 000
	15	Catalyst Unloading	28 000
	2	High Pressure Heavy Flush Oil	1 703 000
		Equipment Cost	$ 16 955 000
		Installation at 12%	2 035 000
		Total Installed Cost	$ 18 990 000

KILBORN **EQUIPMENT** 2-55

CLIENT. Western Canada Coal Liquefaction Study

EQUIPMENT TYPE. Fired Heaters

PROJ. Nº 3349 DIV. NO. 15

AREA 300 - Coal Liquefaction

SHEET Nº ____ CONTº ON ____

REV Nº

DATE REVISED: ____

ISSUE DATE February 1982

EQUIPMENT Nº		DESCRIPTION	ESTIMATED PRICE
Item No.	Quantity		
	5	Slurry Feed Heater	$ 15 365 000
	5	Hydrogen Heater	7 711 000
		Total Installed Cost	$ 23 076 000

KILBORN EQUIPMENT

CLIENT. Western Canada Coal Liquefaction Study
EQUIPMENT TYPE. Vessels
PROJ. Nº. 3349 DIV. NO. 15
AREA 300 - Coal Liquefaction
SHEET Nº CONT'D ON
REV Nº DATE REVISED:
ISSUE DATE February 1982

EQUIPMENT Nº Item No.	Quantity	DESCRIPTION	ESTIMATED PRICE
	5	Slurry Preparation Tank	462 000
	5	Condensate Collector	27 000
	5	Slurry Holding Tank	791 000
	5	High Pressure Steam	323 000
	5	Reactor Effluent Steam	15 595 000
	5	High Pressure Slurry Letdown Drum	9 120 000
	5	Low Pressure Slurry Letdown Drum	471 000
	5	Low Pressure Condensate reclaver	35 000
	5	Low Pressure Condensate Wash Water Separator	22 000
	5	Reactor Effluent Vapour Separator	2 405 000
	5	High Pressure Condensate Receiver	2 490 000
	5	Low Pressure Steam Drum	220 000
	5	Reactor Effluent Vapour Wash Water Separator	2 085 000
	5	High Pressure Condensate	415 000
	5	Compressor Suction Knockout Drum	727 000
	5	Recycle Slurry Oil Flash Drum	120 000
	5	Hydroclone Overflow Surge Drum	160 000
	5	Hydroclone Underflow Surge Drum	125 000
	5	Water Degassing Drum	155 000
	5	Catalyst Fill Hopper	33 000
	5	Catalyst Addition Drum	235 000
	5	Catalyst Dump Tank	39 000
	5	Catalyst Withdrawal Drum	965 000
	5	Catalyst Liquid Separator	1 440 000
		Equipment Cost	$ 38 460 000
		Installation at 10%	3 846 000
		Total Installed Cost	$ 42 306 000

KILBORN **EQUIPMENT** 2-57

CLIENT. Western Canada Coal Liquefaction Study

EQUIPMENT TYPE. Reactors

PROJ.Nº. 3349 DIV.NO. 15

AREA 300 - Coal Liquefaction

SHEET Nº CONTº ON

REV Nº DATE REVISED: ISSUE DATE February 1982

EQUIPMENT Nº		DESCRIPTION	ESTIMATED PRICE
Item No.	Quantity		
	5	H-COAL Reactor	$ 52 500 000
		Equipment Cost	52 500 000
		Installation at 12%	6 300 000
		Total Installed Cost	$ 58 800 000

KILBORN **EQUIPMENT** 2-58

CLIENT. Western Canada Coal Liquefaction Study

EQUIPMENT TYPE. Package Units

PROJ.Nº 3349 DIV. NO. 15

AREA 300 - Coal Liquefaction

SHEET Nº ____ CONTº ON ____

REV Nº

DATE REVISED: ____

ISSUE DATE February 1982

EQUIPMENT Nº		DESCRIPTION	ESTIMATED PRICE
Item No.	Quantity		
	1	Amine Unit	$ 3 490 000
	1	Cryogenic Unit	2 617 000
	5	Hydroclone Assembly	220 000
	5	Ball Mill Package	3 210 000
		Total Installed Cost	$ 9 537 000

KILBORN EQUIPMENT

CLIENT Western Canada Coal Liquefaction Study

EQUIPMENT TYPE Pumps

PROJ. Nº 3349 DIV. NO. 15

AREA 400 - Atmospheric Vacuum Distillation

SHEET Nº CONTº ON

REV Nº DATE REVISED: ISSUE DATE February 1982

EQUIPMENT Nº Item No.	Quantity	DESCRIPTION	ESTIMATED PRICE
	2	Fractionator Upper Pumparound	$ 116 000
	2	Fractionator Lower Pumparound	36 000
	2	Naphtha Feed Pump	29 000
	2	HW Distillate Pump	29 000
	2	Mid Distillate Pump	29 000
	2	Steam Stripper Bottoms	72 000
	2	Fractionator Feed	36 000
	2	Sour Water	40 000
	2	Slurry Oil Recycle	160 000
	2	Vacuum Tower Pumparound	44 000
	2	Vacuum Bottoms	174 000
	2	Naphtha Stabilizer	22 000
		Total Equipment Cost	787 000
		Installation at 12%	94 000
		Total Installed Cost	$ 881 000

KILBORN **EQUIPMENT** 2-60

CLIENT Western Canada Coal
Liquefaction Study

EQUIPMENT TYPE Vessels

PROJ. Nº 3349 DIV. NO. 15

AREA 400 - Atmospheric & Vacuum Distillation

SHEET Nº ____ CONT'D ON ____

REV Nº

DATE REVISED: ____

ISSUE DATE February 1982

EQUIPMENT Nº		DESCRIPTION	ESTIMATED PRICE
Item No.	Quantity		
	1	Fractionator Overhead Drum	$ 68 000
	1	Low Pressure Wash Water Separator	71 000
	1	Naphtha Stabilizer Overhead Drum	23 000
		Equipment Cost	$ 162 000
		Installation at 10%	16 000
		Total Installed Cost	$ 178 000

KILBORN **EQUIPMENT** 2-61

CLIENT. Western Canada Coal Liquefaction Study

EQUIPMENT TYPE. Towers

PROJ. Nº. 3349 DIV. NO. 15

AREA 400 - Atmospheric & Vacuum Distillation

SHEET Nº ____ CONT'D ON ____

REV Nº

DATE REVISED: ____

ISSUE DATE February 1982

EQUIPMENT Nº		DESCRIPTION	ESTIMATED PRICE
Item No.	Quantity		
	1	Fractionator	$ 1 657 000
	1	Steam Stripper	225 000
	1	Vacuum Tower	560 000
	1	Naphtha Stabilizer	173 000
	1	Side Stream Stripper	99 000
		Equipment Cost	$ 2 714 000
		Installation at 12%	326 000
		Total Installed Cost	$ 3 040 000

KILBORN **EQUIPMENT** 2-62

CLIENT Western Canada Coal Liquefaction Study

EQUIPMENT TYPE Heat Exchangers

PROJ. Nº 3349 DIV. NO. 15

AREA 400 - Atmospheric & Vacuum Distillation

SHEET Nº CONT'D ON

REV Nº

DATE REVISED:

ISSUE DATE February 1982

EQUIPMENT Nº		DESCRIPTION	ESTIMATED PRICE
Item No.	Quantity		
	1	Fractionator Upper Pumparound Cooler	$ 883 000
	1	Fractionator Lower Pumparound Cooler	128 000
	1	Fractionator Overhead Air Cooler	581 000
	1	Fractionator Overhead Final Cooler	100 000
	1	Hvy Distillate Cooler	119 000
	1	Mid Distillate Cooler	119 000
	1	Feed Cooler/Waste Feed Heater	177 000
	1	Feed Final Cooler	233 000
	1	Vacuum Tower Pumparound	143 000
	1	Stabilizer Feed/Bottoms Exchanger	29 000
	1	Stabilizer Reboiler	41 000
	1	Stabilizer Overhead Condenser	29 000
	1	Naphtha Prod. Cooler	100 000
		Equipment Cost	$ 2 682 000
		Installation at 8%	215 000
		Total Installed Cost	$ 2 897 000

KILBORN EQUIPMENT 2-63

CLIENT. Western Canada Coal Liquefaction Study

EQUIPMENT TYPE. Fired Heaters

PROJ. Nº. 3349 DIV. NO. 15

AREA 400 - Atmospheric & Vacuum Distillation

SHEET Nº CONTº ON

REV Nº

DATE REVISED:

ISSUE DATE February 1982

EQUIPMENT Nº		DESCRIPTION	ESTIMATED PRICE
Item No.	Quantity		
	1	Fractionator Feed Heater	$ 6 063 000
		Total Installed Cost	$ 6 063 000

KILBORN **EQUIPMENT** 2-64

CLIENT	Western Canada Coal Liquefaction Study
EQUIPMENT TYPE	Package System
PROJ. Nº	3349
DIV. NO.	15
AREA	400 - Atmospheric & Vacuum Distillation
SHEET Nº	
CONT'D ON	
REV. Nº	
DATE REVISED:	
ISSUE DATE	February 1982

EQUIPMENT Nº		DESCRIPTION	ESTIMATED PRICE
Item No.	Quantity		
	1	Residue Slating	$ 1 751 000
	1	Ejector Assembly	55 000
		Total Installed Cost	$ 1 806 000

Preliminary Economic Analysis

3.0 PRELIMINARY ECONOMIC ANALYSIS

3.1 GENERAL

The economic evaluation of the Base Case coal liquefaction complex was facilitated by the development of a cash flow model to simulate the before and after tax cash flow. These simulations were used to calculate discounted cash flow (DCF) rates of return to 100 percent equity.

Sensitivity analyses were conducted on the Base Case at two different coal prices to evaluate alternative plant configurations and monitor the effect of certain parameters on the rate of return. Parameters used in the financial model for specific cases are tabulated in Table 3.1.

3.2 BASE CASE ANALYSIS

The project profitability for the Base Case coal liquefaction complex with coal at $80.00 per tonne, defined as the DCF rate of return to 100 percent equity, was 4.6 percent. A financial analysis summary prepared by the economic evaluation program is included as Figure 3.4 of this report.

For coal at $50.00 per tonne, Base Case DCF rate of return is 10.2 percent. A financial analysis summary of this case is presented as Figure 3.5.

3.3 SENSITIVITY ANALYSIS

Sensitivity Analysis is used to examine methods of improving project profitability and testing the sensitivity of the project profitability to variations in key parameters. In addition, four key parameters were examined for their sensitivity to the project profitability: coal liquefaction yield, capital cost, operating cost and gross revenue. Uncertainty as to the reliability of the 40 percent distillate yield selected led to the need for a specific economic case to determine the project profitability at higher distillate yields.

Total Base Case operating costs, gross revenues and capital costs were varied from minus 50 percent to plus 50 percent of their base values. Gross revenues were varied upward (increased) until a DCF rate of return of 20 percent was reached. Project DCF rates of return for these variations are tabulated in Tables 3.2, 3.3 and 3.4. A plot of the effect of these variations on the rates of return is provided in Figure 3.1. Similar variations were determined for $50.00 per tonne coal. DCF rates of return are tabulated in Tables 3.2, 3.3 and 3.4 and plotted in Figure 3.2.

By analyzing the total operating cost and gross revenue variations the effect of specific components such as raw material and product pricing on the rate of return can be determined. The effect of coal price is shown in Figure 3.3.

Alternative Cases

Case 1

Based on the Base Case coal liquefaction complex but with 50 percent distillate yield and increases to the process facilities to accommodate the increased throughput. Project DCF rate of return is 8.4 percent. A financial analysis summary of this case is available as Figure 3.6 of this report.

Three alternative plant configurations were developed to maximize the utilization of waste by-products for conversion to saleable products. Briefly these configurations were:

Case 2

Base Case process facilities with no power generation. Excess residue and fuel gas are gasified and converted to finished gasoline using the Mobil MTG process. Plant power is purchased.

Project DCF rate of return is 7.6 percent.

Case 3

Base Case process facilities with no power generation. Coal wash plant tailings, excess residue and fuel gas are gasified and converted to finished gasoline using the Mobil MTG process. Plant power is purchased.

Project DCF rate of return is 8.1 percent.

Case 4

Base Case process facilities with no power generation. Coal washplant tailings and excess residue are gasified and converted to finished gasoline using the Mobil MTG process. Steam reforming of natural gas is used to produce hydrogen. Plant power is purchased.

Project DCF rate of return is 7.7 percent.

Financial analysis summaries for these plant configurations are included as Figure 3.7, 3.8 and 3.9 of this report.

One additional alternative was developed based on the concept of providing a simple-to-operate coal liquefaction facility.

Case 5

Base Case process facilities with inplant power generation. Steam reforming of natural gas is used to produce hydrogen. Liquefaction plant solid residue, in excess of the quantity needed for steam and power generation, is discarded as a zero value/zero cost material.

Project cash flow is negative as annual operating costs exceed plant revenue. No economic analysis is provided.

Special Cases

Three special sensitivity analysis cases were developed to evaluate special tax situations and larger syncrude discounts. Details of these cases follow:

Corporate Tax Calculations

In this special situation, the Base Case was evaluated on the basis of a corporate tax calculation, rather than a stand-alone project basis, wherein all capital cost allowances and other tax credits were deducted against corporate income in the years that they were earned. In this manner, more rapid write-off of capital equipment can be realized than on a stand-alone project basis, resulting in a higher DCF rate of return.

Project DCF rate of return is 6.6 percent. A financial analysis summary is provided as Figure 3.10 of this report.

Syncrude Discount, $10.00/Barrel

In this special case, a $10.00/barrel discount (in 1982 dollars) was applied to the Base Case syncrude product, rather than the

$3.00/barrel discount used in the Base Case analysis, to evaluate project sensitivity to product price. This is an important parameter as quality of the coal-derived syncrude is not totally defined and its true value to a refiner is not certain.

Project cash flow is negative as annual operating costs marginally exceed plant revenue, over the life of the project. No economic analysis is provided.

Syncrude Discount, $5.00/Barrel

In this special situation, the Base Case was evaluated with a $5.00/barrel discount (in 1982 dollars) applied to the syncrude product. In addition, coal price was set at $60.00 per tonne and operating cost and revenue escalation factors were set different from the Base Case analysis. Annual operating costs, including coal, were assigned zero real growth (i.e. keeping pace with inflation only), while product prices were assigned a one percent real escalation (inflation plus one percent).

Project DCF rate of return is 9.6 percent.

A financial analysis summary is provided as Figure 3.11 of this report.

TABLE 3.1

FINANCIAL ANALYSIS PARAMETERS

				Alternative		
	Base Case	Case 1	Case 2	Case 3	Case 4	Case 5
Capital Cost $ billion	2.020	2.062	2.520	2.630	3.039	2.087
Gross Revenues $ million	437.50	494.05	562.71	599.18	677.87	420.11
Operating Costs $ million	393.08	391.73	452.67	472.44	584.64	494.05
Working Capital $ million	55.49	61.14	69.70	73.36	80.90	55.45
Equity Financing level, percent	100	100	100	100	100	100
Construction Period, years	7	7	7	7	7	7
Plant Production life, years	20	20	20	20	20	20
Operating days per year	310	310	310	310	310	310
Coal, $ per tonne	80	80	80	80	80	80
Electricity, $ per megawatt hour	22.60	22.60	22.60	22.60	22.60	22.60
Natural gas, $ per 1000 SCF	4.00	4.00	4.00	4.00	4.00	4.00

Escalation	9 to 11 percent per year as per Section 1.3 of this volume.
Costs	First quarter 1982 Canadian dollars.

TABLE 3.2

SENSITIVITY OF THE DCF RATE OF RETURN TO OPERATING COST VARIATIONS

OPERATING COST PERCENTAGE	BASE CASE DCF RATE OF RETURN (PERCENT)	
	$80/tonne Coal	$50/tonne Coal
50%	13.58	16.15
75%	10.51	13.86
100%	4.61	10.16
118%	0 (Cal.)	-
125%	NEG.	5.82
144%	-	0 (Cal.)
150%	-	NEG

TABLE 3.3

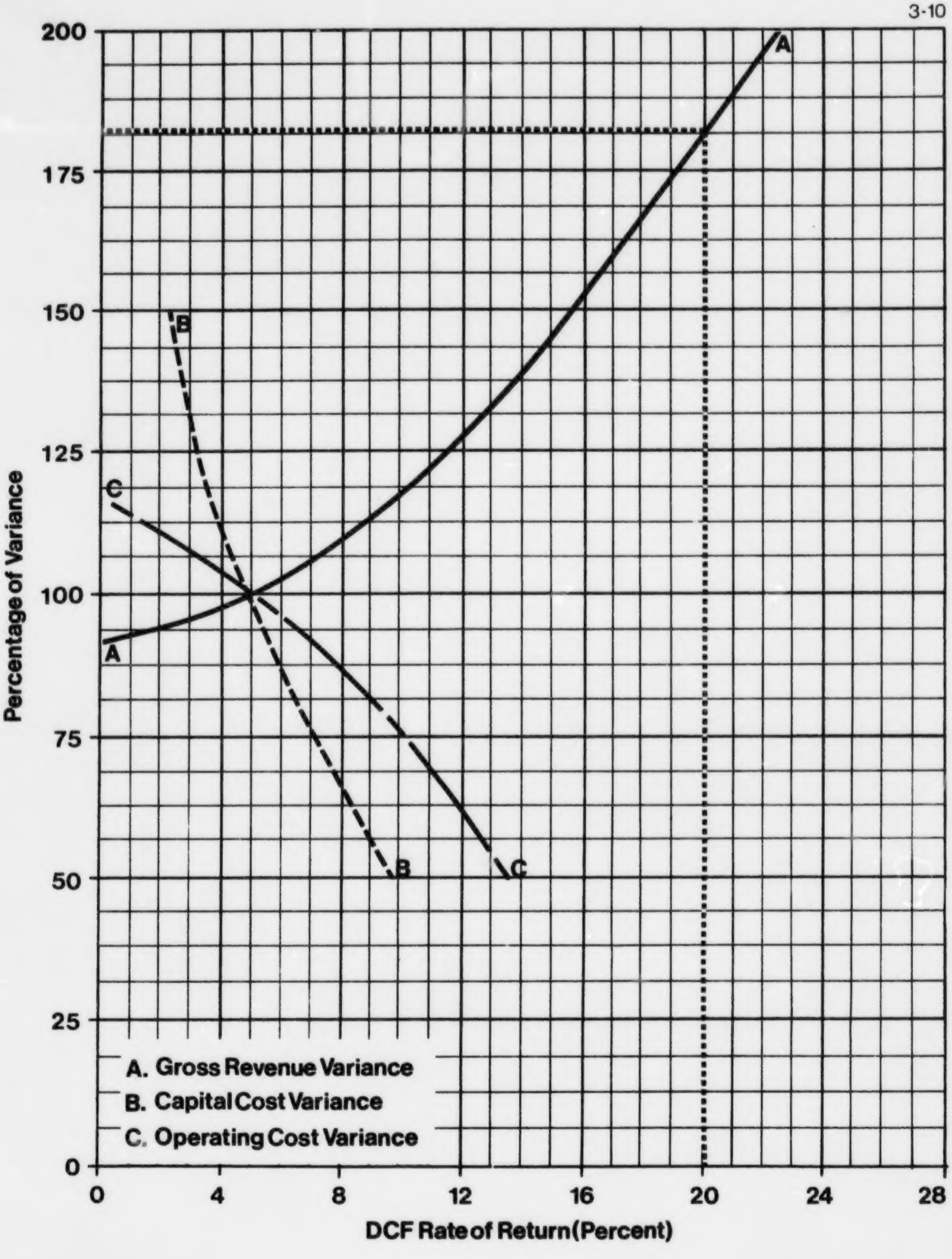
SENSITIVITY OF THE DCF RATE OF RETURN TO GROSS REVENUE VARIATIONS

GROSS REVENUE PERCENTAGE	BASE CASE DCF RATE OF RETURN (PERCENT)	
	$80/tonne Coal	$50/tonne Coal
50%	-	NEG.
71%	-	0 (Cal.)
75%	NEG.	2.33
89%	0 (Cal.)	-
100% (Base)	4.61	10.16
125%	11.41	14.96
150%	15.87	18.63
175%	19.36	21.66
200%	22.27	24.28

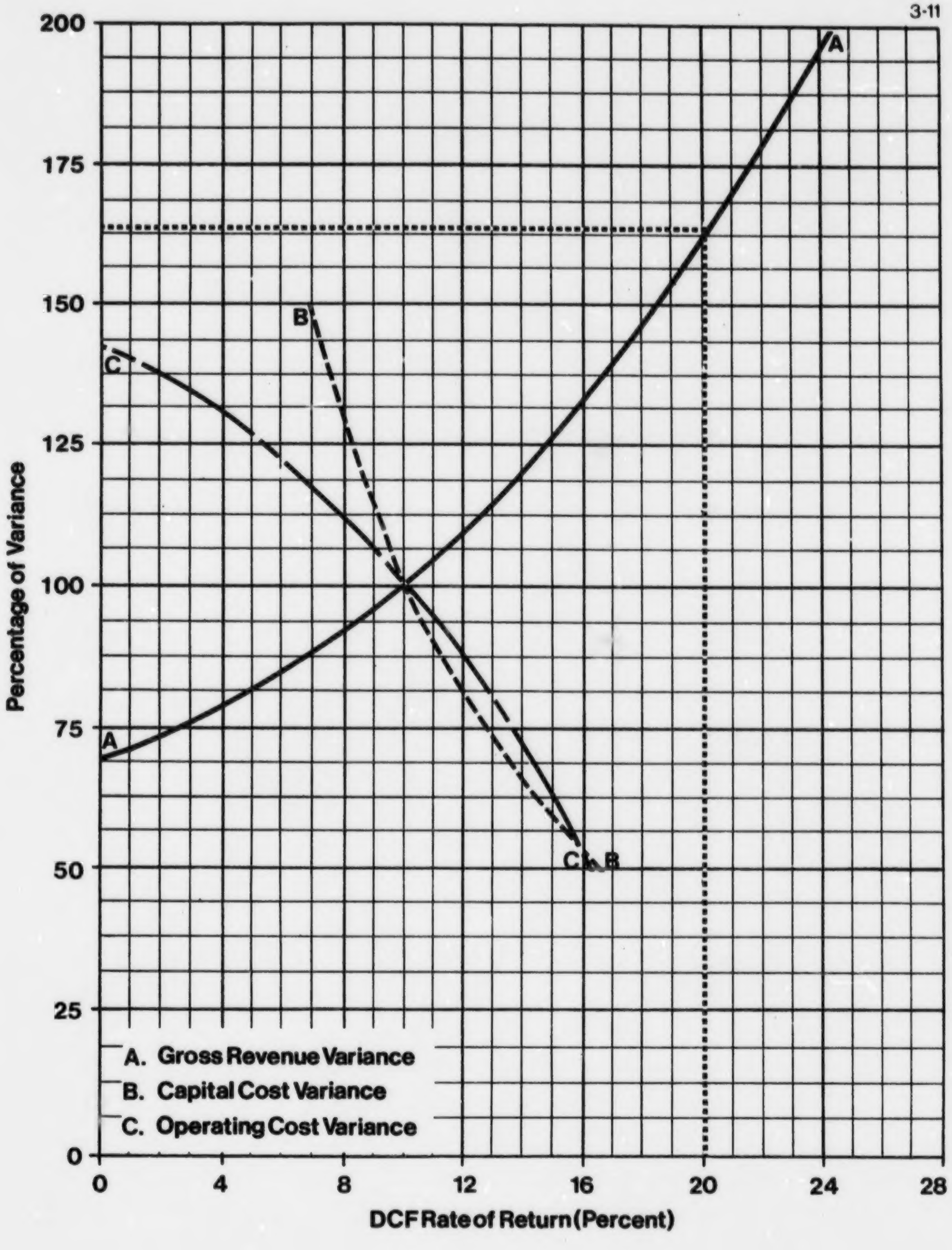
TABLE 3.4

SENSITIVITY OF THE DCF RATE OF RETURN TO CAPITAL COST VARIATIONS

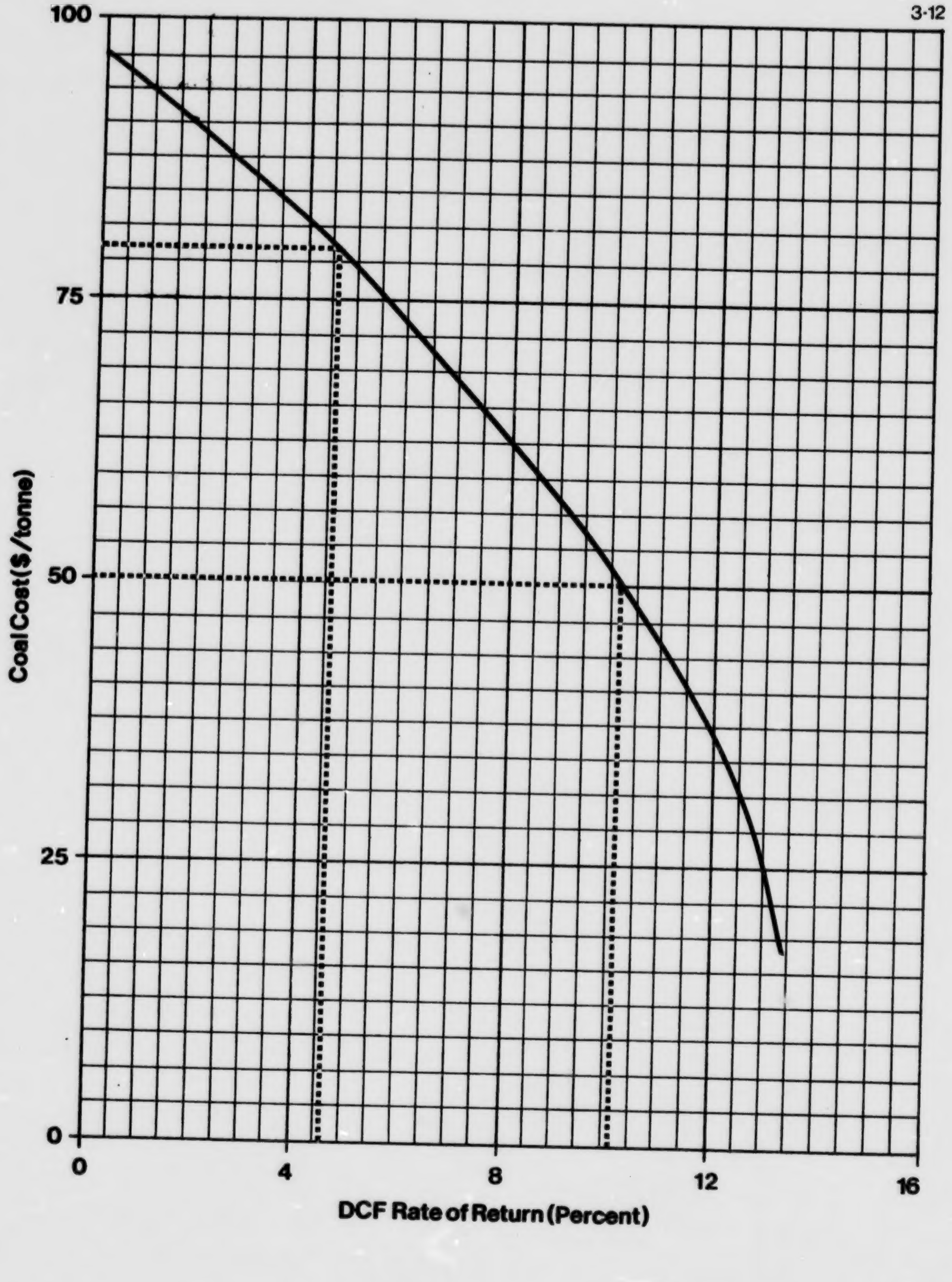
CAPITAL COST PERCENT	BASE CASE DCF RATE OF RETURN (PERCENT)	
	$80/tonne Coal	$50/tonne Coal
50%	9.52	16.54
75%	6.53	12.65
100%	4.61	10.16
125%	3.25	8.39
150%	2.24	6.99



Effect of Gross Revenue, Capital Cost & Operating Cost Variances on Base Case DCF Rate of Return ($80.00/tonne Coal)
Figure 3.1



Effect of Gross Revenue, Capital Cost & Operating Cost Variances on Base Case DCF Rate of Return ($50.00/tonne Coal)
Figure 3.2



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Effect of Coal Cost on
Base Case DCF Rate of Return
Figure 3.3

3.4 ADDENDUM TO ECONOMIC EVALUATION

Based on a revised forecast of new oil reference prices for the period from January 1984 to July 1986 issued by the Province of Alberta in an April 13, 1982 Announcement, discounted cash flow (DCF) calculations were repeated for the Base Case coal liquefaction facility and Alternate Cases 1-5. The revised forecast of new oil prices was issued to reflect recent changes in projected World Oil prices and the provision in the September 1981 Oil Pricing Agreement that the new oil reference price cannot exceed World Oil price.

A comparison of new oil reference prices for the period between January 1984 and July 1986 follows:

Date	September 1981 Agreement	April 13, 1982 Announcement
January 1984	$60.18	$48.86
July 1984	63.48	51.10
January 1985	66.83	53.30
July 1985	70.23	55.75
January 1986	74.08	58.15
July 1986	77.48	60.85

Pricing policy up to January 1984 is not clear at this time but does not affect this economic evaluation as plant startup is not until 1989. Product prices beyond 1986 are escalated at a combined rate of nine percent (seven percent inflation and two percent real escalation), as in the original economic evaluation.

As was the case in the original economic evaluation, the value of the H-Coal syncrude is discounted to reflect its hydrogen deficiency and low quality as compared to oil sands quality syncrude. The July 1986 H-Coal syncrude price, based on the April 13, 1982 Announcement, is $56.85 per barrel, compared to $72.40 per barrel in the original evaluation.

To illustrate the effect of the April 13, 1982 Announcement new oil price forecasts on current dollar (1982) project revenue, a hypothetical 1982 syncrude price was calculated by discounting the $56.85 per barrel 1986 price to 1982 using the same compounded escalation factor used in the original economic evaluation (i.e. $72.40 per barrel syncrude in 1986 divided by $42.90 per barrel in 1982 or 1.687). This gives a current syncrude price of $33.70 per barrel. On this basis, gross revenue for the Base Case and Alternative Cases are as follows, with the original values being given for comparison:

	Gross Revenue, $1000	
	April 13, 1982 Announcement	September 1981 Agreement
Base Case	349 630	437 500
Alternative Cases		
Case 1	393 700	494 050
Case 2	444 300	562 710
Case 3	472 960	599 180
Case 4	534 770	677 870
Case 5	332 250	420 113

Project DCF rates of return for the Base Case and Alternate Cases 1-5 are summarized in Table 3.5 for the revised oil pricing forecast, at both $80.00 per tonne and $50.00 per tonne coal. All DCF rates of return are negative or zero for the cases with coal at $80.00 per tonne.

Financial analysis summaries are presented for Case 1 at $80.00 per tonne coal and for the Base Case and Cases 1-4 at $50.00 per tonne coal, as Figures 3.12, 3.13, 3.14, 3.15, 3.16 and 3.17 respectively.

TABLE 3.5

COMPARISON OF DCF RATES OF RETURN FOR DIFFERENT OIL PRICING FORECASTS

	DCF Rate of Return (Percent)			
Case	April 13, 1982 Announcement		September 1981 Agreement	
	$80/tonne Coal	$50/tonne Coal	$80/tonne Coal	$50/tonne Coal
Base Case	Neg.	4.33	4.61	10.16
Case 1	0.01	7.74	8.38	12.7 (Cal.)
Case 2	Neg.	6.07	7.57	
Case 3	Neg.	6.35	8.05	
Case 4	Neg.	4.78	7.70	
Case 5	Neg.	Neg.	Neg.	

```
                                                                                                                                                         ($1000)

                                     1982     1983     1984     1985     1986     1987     1988     1989     1990     1991     1992     1993     1994     1995     1996     1997     1998     1999     2000     2001     2002     2003     2004     2005     2006     2007     2008      TOTAL
                                     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----      =====
**********************
CASH FLOW CALCULATIONS
**********************
+ REVENUE                               0        0        0        0        0        0        0   382213   729071  1021741  1237441  1348811  1470204  1602523  1746750  1903957  2075313  2262091  2465680  2687591  2929474  3193127  3480508  3793754  4135192  4507359  4913021   47885820
- OPERATING COSTS                       0        0        0        0        0        0        0   382213   663117   860674  1013277  1104472  1203874  1312223  1430323  1559052  1699367  1852310  2019018  2200730  2398795  2614687  2850009  3106509  3386095  3690844  4023020   39370608
- FEDERAL INCOME TAX                    0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0   118055   190704   207915   226672   247116   269397   293682   320150    1873691
- PROVINCIAL INCOME TAX                 0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    29514    47676    51979    56668    61779    67349    73420    80038     468423
- INTEREST PAID FROM REVENUE            0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0          0
- DEBT REPAYMENTS                       0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0          0
- CAPITAL COSTS FROM REVENUE            0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0          0
- EQUITY CAPITAL COSTS                  0        0   130664   427272   776211  1015284   922216   162534        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    3434181
------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
= CASH FLOW                             0        0  -130664  -427272  -776211 -1015284  -922216  -162534    65954   161067   224164   244339   266330   290299   316426   344905   375946   409781   446662   339293   292299   318546   347159   378349   412350   449413   489814    2738916
  ACCUMULATED                           0        0  -130664  -557936 -1334147 -2349430 -3271646 -3434181 -3368227 -3207160 -2982996 -2738657 -2472327 -2182027 -1865601 -1520696 -1144750  -734968  -288307    50986   343285   661832  1008991  1387340  1799693  2249103  2738916

DCF RATE OF RETURN (%)               4.61
PAYOUT PERIOD (YEARS)               12.85
EQUITY RATIO (%)                   100.00


*******************
COMPONENT INFLATION
*******************
PRICE         RATE                   5.00    11.00    11.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00
              FACTOR                1.050    1.166    1.294    1.410    1.537    1.675    1.826    1.991    2.170    2.365    2.578    2.810    3.063    3.338    3.639    3.966    4.323    4.712    5.136    5.599    6.103    6.652    7.250    7.903    8.614    9.390   10.235

CAPITAL       RATE                   5.00    11.00    11.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00
              FACTOR                1.050    1.166    1.294    1.410    1.537    1.675    1.826    1.991    2.170    2.365    2.578    2.810    3.063    3.338    3.639    3.966    4.323    4.712    5.136    5.599    6.103    6.652    7.250    7.903    8.614    9.390   10.235

OPERATING     RATE                   5.00    11.00    11.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00
              FACTOR                1.050    1.166    1.294    1.410    1.537    1.675    1.826    1.991    2.170    2.365    2.578    2.810    3.063    3.338    3.639    3.966    4.323    4.712    5.136    5.599    6.103    6.652    7.250    7.903    8.614    9.390   10.235


********************
CAPITAL EXPENDITURES
********************
PROCESS FACILITIES                      0        0    86000   180000   300000   350000   264000        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    1180000
BUILDINGS + SERVICES                    0        0        0        0        0    10000    10000        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      20000
INDIRECTS                               0        0    10000    63000   105000   126000   116000        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     420000
CONTINGENCY                             0        0     5000    60000   100000   120000   115000        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     400000
WORKING CAPITAL                         0        0        0        0        0        0        0    55490        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      55490
------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
TOTAL CAPITAL COSTS                     0        0   101000   303000   505000   606000   505000    55490        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    2075490
INFLATION                               0        0    29664   124272   271211   409284   417216    54964        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    1306611
CAPITALIZED COSTS                       0        0        0        0        0        0        0    52080        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      52080
------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
TOTAL CAPITAL                           0        0   130664   427272   776211  1015284   922216   162534        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    3434181
ACCUMULATED                             0        0   130664   557936  1334147  2349430  3271646  3434181  3434181  3434181  3434181  3434181  3434181  3434181  3434181  3434181  3434181  3434181  3434181  3434181  3434181  3434181  3434181  3434181  3434181  3434181  3434181


***************
OPERATING COSTS
***************
OPERATING COSTS                         0        0        0        0        0        0        0   218180   305630   363930   393080   393080   393080   393080   393080   393080   393080   393080   393080   393080   393080   393080   393080   393080   393080   393080   393080    7570100
INFLATION                               0        0        0        0        0        0        0   216113   357487   496744   620197   711392   810794   919143  1037243  1165972  1306287  1459230  1625938  1807650  2005715  2221607  2456929  2713429  2993015  3297764  3629940   31852588
------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
TOTAL OPERATING COSTS                   0        0        0        0        0        0        0   434293   663117   860674  1013277  1104472  1203874  1312223  1430323  1559052  1699367  1852310  2019018  2200730  2398795  2614687  2850009  3106509  3386095  3690844  4023020   39422688
CAPITALIZED COSTS                       0        0        0        0        0        0        0    52080        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      52080
------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
TOTAL OPERATING COSTS                   0        0        0        0        0        0        0   382213   663117   860674  1013277  1104472  1203874  1312223  1430323  1559052  1699367  1852310  2019018  2200730  2398795  2614687  2850009  3106509  3386095  3690844  4023020   39370608


*****************
 TAX CALCULATIONS
*****************

FEDERAL INCOME TAX
******************
+ REVENUE                               0        0        0        0        0        0        0   382213   729071  1021741  1237441  1348811  1470204  1602523  1746750  1903957  2075313  2262091  2465680  2687591  2929474  3193127  3480508  3793754  4135192  4507359  4913021   47885820
- OPERATING COSTS                       0        0        0        0        0        0        0   382213   663117   860674  1013277  1104472  1203874  1312223  1430323  1559052  1699367  1852310  2019018  2200730  2398795  2614687  2850009  3106509  3386095  3690844  4023020   39370608
- INTEREST                              0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0          0
- CLASS 3                               0        0        0        0        0        0        0        0     1751     1663     1580     1501     1426     1355     1287     1223     1162     1103     1048      996      946      899      854      811      771      732      695      21802
- CLASS 29 CURRENT                      0        0        0        0        0        0        0        0    64203   159404        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     223607
- CLASS 29 CARRIED                      0        0        0        0        0        0        0        0        0        0   222584   242838   264904   288945   315139   343682   374785   408678   445613   157935        0        0        0        0        0        0        0    3065104
------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
= TAXABLE INCOME                        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0   327930   529733   577541   629646   686433   748326   815783   889306    5204698
  INCOME TAX (36.0%)                    0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0   118055   190704   207915   226672   247116   269397   293682   320150    1873691


PROVINCIAL INCOME TAX
*********************
FEDERAL TAXABLE INCOME                  0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0   327930   529733   577541   629646   686433   748326   815783   889306    5204698
INCOME TAX (9.0 %)                      0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    29514    47676    51979    56668    61779    67349    73420    80038     468423


CAPITAL COST ALLOWANCE
**********************
INPUT
   CLASS 3                              0        0        0        0        0    16754    18262        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      35016
   CLASS 29                             0        0   130664   427272   776211   998530   903954    52080        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    3288711
   NO ALLOWANCE                         0        0        0        0        0        0        0   110454        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     110454
------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
   TOTAL INPUT                          0        0   130664   427272   776211  1015284   922216   162534        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    3434181

POOLS BEFORE DEDUCTIONS
   CLASS 3                              0        0        0        0        0    16754    35016    35016    35016    33265    31602    30021    28520    27094    25740    24453    23230    22069    20965    19917    18921    17975    17076    16222    15411    14641    13909
   CLASS 29 CURRENT                     0        0        0        0        0        0        0   822178  1644355   822178        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0
   CLASS 29 CARRIED                     0        0        0        0        0        0        0   822178  2402330  3065104  3065104  2842520  2599682  2334778  2045833  1730694  1387012  1012227   603549   157935        0        0        0        0        0        0        0
```

Base Case
($80.00/tonne Coal)
September, 1981 Oil Pricing Agreement
Figure 3.4

($1000)

CASH FLOW CALCULATIONS	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
+ REVENUE	0	0	0	0	0	0	0	382213	729071	1021741	1237441	1348811	1470204	1602523	1746750	1903957	2075313	2262091	2465680	2687591	2929474	3193127	3480508	3793754	4135182	4507359	4913021	47885828
- OPERATING COSTS	0	0	0	0	0	0	0	360245	521871	662728	773543	843162	919046	1001760	1091919	1190191	1297309	1414066	1541332	1680052	1831257	1996070	2175716	2371531	2584959	2817616	3071201	30145584
- FEDERAL INCOME TAX	0	0	0	0	0	0	0	0	0	0	0	0	0	0	41783	256538	279684	304912	332407	362373	395035	430633	469433	511723	557817	608057	662817	5213211
- PROVINCIAL INCOME TAX	0	0	0	0	0	0	0	0	0	0	0	0	0	0	10446	64134	69921	76228	83102	90593	98759	107658	117358	127931	139454	152014	165704	1303303
- INTEREST PAID FROM REVENUE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
- DEBT REPAYMENTS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
- CAPITAL COSTS FROM REVENUE	0	0	0	0	0	0	0	21967	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	21967
- EQUITY CAPITAL COSTS	0	0	130664	427272	776211	1015284	922216	88487	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	3360133
= CASH FLOW	0	0	-130664	-427272	-776211	-1015284	-922216	-88487	207199	359013	463899	505650	551158	600762	602603	393094	428399	466886	508839	554572	604424	658765	718001	782569	852952	929672	1013298	7841621
ACCUMULATED	0	0	-130664	-557936	-1334147	-2349430	-3271646	-3360133	-3152934	-2793921	-2330022	-1824373	-1273215	-672452	-69850	323244	751643	1218529	1727368	2281940	2886364	3545129	4263130	5045699	5898651	6828323	7841621	

DCF RATE OF RETURN (%) 10.16
PAYOUT PERIOD (YEARS) 8.10
EQUITY RATIO (%) 100.00

COMPONENT INFLATION		1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008
PRICE	RATE	5.00	11.00	11.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00
	FACTOR	1.050	1.166	1.294	1.410	1.537	1.675	1.826	1.991	2.170	2.365	2.578	2.810	3.063	3.338	3.639	3.966	4.323	4.712	5.136	5.599	6.103	6.652	7.250	7.903	8.614	9.390	10.235
CAPITAL	RATE	5.00	11.00	11.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00
	FACTOR	1.050	1.166	1.294	1.410	1.537	1.675	1.826	1.991	2.170	2.365	2.578	2.810	3.063	3.338	3.639	3.966	4.323	4.712	5.136	5.599	6.103	6.652	7.250	7.903	8.614	9.390	10.235
OPERATING	RATE	5.00	11.00	11.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00
	FACTOR	1.050	1.166	1.294	1.410	1.537	1.675	1.826	1.991	2.170	2.365	2.578	2.810	3.063	3.338	3.639	3.966	4.323	4.712	5.136	5.599	6.103	6.652	7.250	7.903	8.614	9.390	10.235

CAPITAL EXPENDITURES	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
PROCESS FACILITIES	0	0	86000	180000	300000	350000	264000	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1180000
BUILDINGS + SERVICES	0	0	0	0	0	10000	10000	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	20000
INDIRECTS	0	0	10000	63000	105000	126000	116000	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	420000
CONTINGENCY	0	0	5000	60000	100000	120000	115000	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	400000
WORKING CAPITAL	0	0	0	0	0	0	0	55490	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	55490
TOTAL CAPITAL COSTS	0	0	101000	303000	505000	606000	505000	55490	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	2075490
INFLATION	0	0	29664	124272	271211	409284	417216	54964	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1306611
CAPITALIZED COSTS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
TOTAL CAPITAL	0	0	130664	427272	776211	1015284	922216	110454	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	3382101
ACCUMULATED	0	0	130664	557936	1334147	2349430	3271646	3382101	3382101	3382101	3382101	3382101	3382101	3382101	3382101	3382101	3382101	3382101	3382101	3382101	3382101	3382101	3382101	3382101	3382101	3382101	3382101	

OPERATING COSTS	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
OPERATING COSTS	0	0	0	0	0	0	0	180980	240530	280230	300080	300080	300080	300080	300080	300080	300080	300080	300080	300080	300080	300080	300080	300080	300080	300080	300080	5803100
INFLATION	0	0	0	0	0	0	0	179265	281341	382498	473463	543082	618966	701680	791839	890111	997229	1113986	1241252	1379972	1531177	1695990	1875636	2071451	2284879	2517536	2771121	24342484
TOTAL OPERATING COSTS	0	0	0	0	0	0	0	360245	521871	662728	773543	843162	919046	1001760	1091919	1190191	1297309	1414066	1541332	1680052	1831257	1996070	2175716	2371531	2584959	2817616	3071201	30145584
CAPITALIZED COSTS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
TOTAL OPERATING COSTS	0	0	0	0	0	0	0	360245	521871	662728	773543	843162	919046	1001760	1091919	1190191	1297309	1414066	1541332	1680052	1831257	1996070	2175716	2371531	2584959	2817616	3071201	30145584

TAX CALCULATIONS

FEDERAL INCOME TAX	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
+ REVENUE	0	0	0	0	0	0	0	382213	729071	1021741	1237441	1348811	1470204	1602523	1746750	1903957	2075313	2262091	2465680	2687591	2929474	3193127	3480508	3793754	4135182	4507359	4913021	47885828
- OPERATING COSTS	0	0	0	0	0	0	0	360245	521871	662728	773543	843162	919046	1001760	1091919	1190191	1297309	1414066	1541332	1680052	1831257	1996070	2175716	2371531	2584959	2817616	3071201	30145584
- INTEREST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
- CLASS 3	0	0	0	0	0	0	0	1751	1663	1580	1501	1426	1355	1287	1223	1162	1103	1048	996	946	899	854	811	771	732	695	661	22463
- CLASS 29 CURRENT	0	0	0	0	0	0	0	20217	205536	357433	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	583186
- CLASS 29 CARRIED	0	0	0	0	0	0	0	0	0	0	462398	504224	549803	599475	537545	0	0	0	0	0	0	0	0	0	0	0	0	2653446
= TAXABLE INCOME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	116063	712604	776901	846977	923352	1006593	1097318	1196203	1303981	1421452	1549491	1689048	1841159	14481142
INCOME TAX (36.0%)	0	0	0	0	0	0	0	0	0	0	0	0	0	0	41783	256538	279684	304912	332407	362373	395035	430633	469433	511723	557817	608057	662817	5213211

PROVINCIAL INCOME TAX	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
FEDERAL TAXABLE INCOME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	116063	712604	776901	846977	923352	1006593	1097318	1196203	1303981	1421452	1549491	1689048	1841159	14481142
INCOME TAX (9.0 %)	0	0	0	0	0	0	0	0	0	0	0	0	0	0	10446	64134	69921	76228	83102	90593	98759	107658	117358	127931	139454	152014	165704	1303303

CAPITAL COST ALLOWANCE	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
INPUT																												
CLASS 3	0	0	0	0	0	16754	18262	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	35016
CLASS 29	0	0	130664	427272	776211	998530	903954	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	3236631
NO ALLOWANCE	0	0	0	0	0	0	0	110454	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	110454
TOTAL INPUT	0	0	130664	427272	776211	1015284	922216	110454	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	3382101
POOLS BEFORE DEDUCTIONS																												
CLASS 3	0	0	0	0	0	16754	35016	35016	33265	31602	30021	28520	27094	25740	24453	23230	22068	20965	19917	18921	17975	17076	16222	15411	14641	13909	13213	
CLASS 29 CURRENT	0	0	0	0	0	0	0	809158	1618315	809158	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	
CLASS 29 CARRIED	0	0	0	0	0	0	0	788941	2201720	2653446	2653446	2191048	1686824	1137021	537545	0	0	0	0	0	0	0	0	0	0	0	0	

Base Case
($50.00/tonne Coal)
September, 1981 Oil Pricing Agreement
Figure 3.5

```
                                                                                                                               ($1000)

                                   1982     1983     1984     1985     1986     1987     1988     1989     1990     1991     1992     1993     1994     1995     1996     1997     1998     1999     2000     2001     2002     2003     2004     2005     2006     2007     2008        TOTAL
                                   ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----        =====
************************
CASH FLOW CALCULATIONS
************************
+ REVENUE                             0        0        0        0        0        0        0   429969   820169  1149407  1472060  1517345  1653906  1802758  1965006  2141857  2334624  2544740  2773766  3023405  3295512  3592108  3915398  4267783  4651894  5070553  5526903     53869153
- OPERATING COSTS                     0        0        0        0        0        0        0   429969   659982   857398  1089784  1100665  1199724  1307700  1425393  1553678  1693509  1845925  2012058  2193143  2390526  2605674  2840184  3095801  3374423  3678121  4009152     39282809
- FEDERAL INCOME TAX                  0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     3280   230383   251176   273838   298536   325454   354793   386769   421622   459609   501012   546140      4052612
- PROVINCIAL INCOME TAX               0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      820    57596    62794    68459    74634    81364    88698    96692   105405   114902   125253   136535      1013153
- INTEREST PAID FROM REVENUE          0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0            0
- DEBT REPAYMENTS                     0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0            0
- CAPITAL COSTS FROM REVENUE          0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0            0
- EQUITY CAPITAL COSTS                0        0   130664   437143   791582  1032037   949608   122969        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      3464003
-----------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
= CASH FLOW                           0        0  -130664  -437143  -791582 -1032037  -949608  -122969   160187   292009   382276   416680   454182   495058   539613   584079   353136   384845   419411   457092   498168   542943   591752   644955   702950   766167   835076      6056576
  ACCUMULATED                         0        0  -130664  -567807 -1359389 -2391426 -3341034 -3464003 -3303816 -3011807 -2629531 -2212851 -1758669 -1263611  -723998  -139919   213216   598061  1017472  1474565  1972732  2515676  3107427  3752383  4455333  5221500  6056576

DCF RATE OF RETURN (%)             8.38
PAYOUT PERIOD (YEARS)              9.40
EQUITY RATIO (%)                 100.00


********************
COMPONENT INFLATION
********************
PRICE        RATE                  5.00    11.00    11.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00
             FACTOR               1.050    1.166    1.294    1.410    1.537    1.675    1.826    1.991    2.170    2.365    2.578    2.810    3.063    3.338    3.639    3.966    4.323    4.712    5.136    5.599    6.103    6.652    7.250    7.903    8.614    9.390   10.235

CAPITAL      RATE                  5.00    11.00    11.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00
             FACTOR               1.050    1.166    1.294    1.410    1.537    1.675    1.826    1.991    2.170    2.365    2.578    2.810    3.063    3.338    3.639    3.966    4.323    4.712    5.136    5.599    6.103    6.652    7.250    7.903    8.614    9.390   10.235

OPERATING    RATE                  5.00    11.00    11.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00
             FACTOR               1.050    1.166    1.294    1.410    1.537    1.675    1.826    1.991    2.170    2.365    2.578    2.810    3.063    3.338    3.639    3.966    4.323    4.712    5.136    5.599    6.103    6.652    7.250    7.903    8.614    9.390   10.235


********************
CAPITAL EXPENDITURES
********************
PROCESS FACILITIES                    0        0    86000   185000   305000   355000   271160        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      1202160
BUILDINGS + SERVICES                  0        0        0        0        0    10000    10000        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        20000
INDIRECTS                             0        0    10000    63000   107000   128000   119740        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0       427740
CONTINGENCY                           0        0     5000    62000   103000   123000   119100        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0       412100
WORKING CAPITAL                       0        0        0        0        0        0        0    61140        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        61140
-----------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
TOTAL CAPITAL COSTS                   0        0   101000   310000   515000   616000   520000    61140        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      2123140
INFLATION                             0        0    29664   127143   276582   416037   429608    60561        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      1339595
CAPITALIZED COSTS                     0        0        0        0        0        0        0     1268        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0         1268
-----------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
TOTAL CAPITAL                         0        0   130664   437143   791582  1032037   949608   122969        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      3464003
ACCUMULATED                           0        0   130664   567807  1359389  2391426  3341034  3464003  3464003  3464003  3464003  3464003  3464003  3464003  3464003  3464003  3464003  3464003  3464003  3464003  3464003  3464003  3464003  3464003  3464003  3464003  3464003


***************
OPERATING COSTS
***************
OPERATING COSTS                       0        0        0        0        0        0        0   216645   304185   362545   391725   391725   391725   391725   391725   391725   391725   391725   391725   391725   391725   391725   391725   391725   391725   391725   391725      7542700
INFLATION                             0        0        0        0        0        0        0   214592   355797   494853   618059   708940   807999   915975  1033668  1161953  1301784  1454200  1620333  1801418  1998801  2213949  2448459  2704076  2982698  3286396  3617427     31741377
-----------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
TOTAL OPERATING COSTS                 0        0        0        0        0        0        0   431237   659982   857398  1089784  1100665  1199724  1307700  1425393  1553678  1693509  1845925  2012058  2193143  2390526  2605674  2840184  3095801  3374423  3678121  4009152     39284077
CAPITALIZED COSTS                     0        0        0        0        0        0        0     1268        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0         1268
-----------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
TOTAL OPERATING COSTS                 0        0        0        0        0        0        0   429969   659982   857398  1089784  1100665  1199724  1307700  1425393  1553678  1693509  1845925  2012058  2193143  2390526  2605674  2840184  3095801  3374423  3678121  4009152     39282809


*****************
 TAX CALCULATIONS
*****************

FEDERAL INCOME TAX
******************
+ REVENUE                             0        0        0        0        0        0        0   429969   820169  1149407  1472060  1517345  1653906  1802758  1965006  2141857  2334624  2544740  2773766  3023405  3295512  3592108  3915398  4267783  4651894  5070553  5526903     53869153
- OPERATING COSTS                     0        0        0        0        0        0        0   429969   659982   857398  1089784  1100665  1199724  1307700  1425393  1553678  1693509  1845925  2012058  2193143  2390526  2605674  2840184  3095801  3374423  3678121  4009152     39282809
- INTEREST                            0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0            0
- CLASS 3                             0        0        0        0        0        0        0        0     1751     1663     1580     1501     1426     1355     1287     1223     1162     1103     1048      996      946      899      854      811      771      732      695        21802
- CLASS 29 CURRENT                    0        0        0        0        0        0        0        0   158436   290346        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0       448782
- CLASS 29 CARRIED                    0        0        0        0        0        0        0        0        0        0   380696   415179   452756   493703   538326   577844        0        0        0        0        0        0        0        0        0        0        0      2858505
-----------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
= TAXABLE INCOME                      0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     9112   639953   697711   760660   829266   904040   985536  1074360  1171171  1276698  1391700  1517056     11257255
  INCOME TAX (36.0%)                  0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     3280   230383   251176   273838   298536   325454   354793   386769   421622   459609   501012   546140      4052612


PROVINCIAL INCOME TAX
*********************
FEDERAL TAXABLE INCOME                0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     9112   639953   697711   760660   829266   904040   985536  1074360  1171171  1276698  1391700  1517056     11257255
INCOME TAX (9.0 %)                    0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      820    57596    62794    68459    74634    81364    88698    96692   105405   114902   125253   136535      1013153

CAPITAL COST ALLOWANCE
**********************
INPUT
   CLASS 3                            0        0        0        0        0    16754    18262        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        35016
   CLASS 29                           0        0   130664   437143   791582  1015283   931347     1268        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      3307287
   NO ALLOWANCE                       0        0        0        0        0        0        0   121701        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0       121701
-----------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
   TOTAL INPUT                        0        0   130664   437143   791582  1032037   949608   122969        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      3464003

POOLS BEFORE DEDUCTIONS
   CLASS 3                            0        0        0        0        0    16754    35016    35016    35016    33265    31602    30021    28520    27094    25740    24453    23230    22069    20965    19917    18921    17975    17076    16222    15411    14641    13909
   CLASS 29 CURRENT                   0        0        0        0        0        0        0   826827  1653643   826822        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0
   CLASS 29 CARRIED                   0        0        0        0        0        0        0   826822  2322929  2858505  2858505  2477809  2062630  1609874  1116171   577844        0        0        0        0        0        0        0        0        0        0        0
```

Alternate Case 1
($80.00/tonne Coal)
September, 1981 Oil Pricing
Agreement
Figure 3.6

($1000)

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
CASH FLOW CALCULATIONS																												
+ REVENUE	0	0	0	0	0	0	0	491944	938383	1315079	1592707	1736051	1892295	2062602	2248235	2450577	2671129	2911531	3173569	3459190	3770517	4109863	4479751	4882929	5322392	5801408	6323534	61633688
- OPERATING COSTS	0	0	0	0	0	0	0	491944	790455	1000891	1166887	1271907	1386379	1511153	1647157	1795401	1956987	2133116	2325096	2534355	2762447	3011067	3282063	3577449	3899419	4250367	4632900	45427441
- FEDERAL INCOME TAX	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	79546	279753	304997	332510	362497	395178	430799	469622	511937	558058	608328	4333227
- PROVINCIAL INCOME TAX	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	19887	69938	76249	83128	90624	98795	107700	117406	127984	139515	152082	1083307
- INTEREST PAID FROM REVENUE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
- DEBT REPAYMENTS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
- CAPITAL COSTS FROM REVENUE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
- EQUITY CAPITAL COSTS	0	0	139720	479447	884880	1206278	1417656	196091	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	4324072
= CASH FLOW	0	0	-139720	-479447	-884880	-1206278	-1417656	-196091	147928	314187	425820	464143	505916	551449	601077	655176	614709	428724	467226	509197	554949	604823	659189	718452	783051	853468	930224	6465642
ACCUMULATED	0	0	-139720	-619167	-1504047	-2710325	-4127981	-4324072	-4176143	-3861956	-3436136	-2971993	-2466076	-1914627	-1313548	-658372	-43663	385061	852287	1361484	1916434	2521257	3180446	3898898	4681950	5535417	6465642	
DCF RATE OF RETURN (%)	7.57																											
PAYOUT PERIOD (YEARS)	10.10																											
EQUITY RATIO (%)	100.00																											
COMPONENT INFLATION																												
PRICE RATE	5.00	11.00	11.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	
FACTOR	1.050	1.166	1.294	1.410	1.537	1.675	1.826	1.991	2.170	2.365	2.578	2.810	3.063	3.338	3.639	3.966	4.323	4.712	5.136	5.599	6.103	6.652	7.250	7.903	8.614	9.390	10.235	
CAPITAL RATE	5.00	11.00	11.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	
FACTOR	1.050	1.166	1.294	1.410	1.537	1.675	1.826	1.991	2.170	2.365	2.578	2.810	3.063	3.338	3.639	3.966	4.323	4.712	5.136	5.599	6.103	6.652	7.250	7.903	8.614	9.390	10.235	
OPERATING RATE	5.00	11.00	11.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	
FACTOR	1.050	1.166	1.294	1.410	1.537	1.675	1.826	1.991	2.170	2.365	2.578	2.810	3.063	3.338	3.639	3.966	4.323	4.712	5.136	5.599	6.103	6.652	7.250	7.903	8.614	9.390	10.235	
CAPITAL EXPENDITURES																												
PROCESS FACILITIES	0	0	90000	200000	320000	400000	459700	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1469700
BUILDINGS + SERVICES	0	0	0	0	0	12000	12000	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	24000
INDIRECTS	0	0	12000	70000	130700	157000	153100	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	522800
CONTINGENCY	0	0	6000	70000	125000	151000	151500	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	503500
WORKING CAPITAL	0	0	0	0	0	0	0	69685	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	69685
TOTAL CAPITAL COSTS	0	0	108000	340000	575700	720000	776300	69685	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	2589685
INFLATION	0	0	31720	139447	309180	486278	641356	69025	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1677006
CAPITALIZED COSTS	0	0	0	0	0	0	0	57381	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	57381
TOTAL CAPITAL	0	0	139720	479447	884880	1206278	1417656	196091	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	4324072
ACCUMULATED	0	0	139720	619167	1504047	2710325	4127981	4324072	4324072	4324072	4324072	4324072	4324072	4324072	4324072	4324072	4324072	4324072	4324072	4324072	4324072	4324072	4324072	4324072	4324072	4324072	4324072	
OPERATING COSTS																												
OPERATING COSTS	0	0	0	0	0	0	0	275970	364320	423220	452670	452670	452670	452670	452670	452670	452670	452670	452670	452670	452670	452670	452670	452670	452670	452670	452670	8758900
INFLATION	0	0	0	0	0	0	0	273355	426135	577671	714217	819237	933709	1058483	1194487	1342731	1504317	1680446	1872426	2081685	2309777	2558397	2829393	3124779	3446749	3797697	4180230	36725922
TOTAL OPERATING COSTS	0	0	0	0	0	0	0	549325	790455	1000891	1166887	1271907	1386379	1511153	1647157	1795401	1956987	2133116	2325096	2534355	2762447	3011067	3282063	3577449	3899419	4250367	4632900	45484822
CAPITALIZED COSTS	0	0	0	0	0	0	0	57381	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	57381
TOTAL OPERATING COSTS	0	0	0	0	0	0	0	491944	790455	1000891	1166887	1271907	1386379	1511153	1647157	1795401	1956987	2133116	2325096	2534355	2762447	3011067	3282063	3577449	3899419	4250367	4632900	45427441
TAX CALCULATIONS																												
FEDERAL INCOME TAX																												
+ REVENUE	0	0	0	0	0	0	0	491944	938383	1315079	1592707	1736051	1892295	2062602	2248235	2450577	2671129	2911531	3173569	3459190	3770517	4109863	4479751	4882929	5322392	5801408	6323534	61633688
- OPERATING COSTS	0	0	0	0	0	0	0	491944	790455	1000891	1166887	1271907	1386379	1511153	1647157	1795401	1956987	2133116	2325096	2534355	2762447	3011067	3282063	3577449	3899419	4250367	4632900	45427441
- INTEREST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
- CLASS 3	0	0	0	0	0	0	0	0	2101	1996	1896	1801	1711	1626	1544	1467	1394	1324	1258	1195	1135	1078	1025	973	925	878	835	26163
- CLASS 29 CURRENT	0	0	0	0	0	0	0	0	145827	312192	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	458019
- CLASS 29 CARRIED	0	0	0	0	0	0	0	0	0	0	423924	462342	504205	549823	599535	653709	491786	0	0	0	0	0	0	0	0	0	0	3685324
= TAXABLE INCOME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	220962	777091	847214	923640	1006935	1097718	1196663	1304507	1422046	1550162	1689800	12036741
INCOME TAX (36.0%)	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	79546	279753	304997	332510	362497	395178	430799	469622	511937	558058	608328	4333227
PROVINCIAL INCOME TAX																												
FEDERAL TAXABLE INCOME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	220962	777091	847214	923640	1006935	1097718	1196663	1304507	1422046	1550162	1689800	12036741
INCOME TAX (9.0 %)	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	19887	69938	76249	83128	90624	98795	107700	117406	127984	139515	152082	1083307
CAPITAL COST ALLOWANCE																												
INPUT																												
CLASS 3	0	0	0	0	0	20185	21834	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	42019
CLASS 29	0	0	139720	479447	884880	1186093	1395822	57381	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	4143343
NO ALLOWANCE	0	0	0	0	0	0	0	138710	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	138710
TOTAL INPUT	0	0	139720	479447	884880	1206278	1417656	196091	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	4324072
POOLS BEFORE DEDUCTIONS																												
CLASS 3	0	0	0	0	0	20185	42019	42019	42019	39918	37922	36026	34224	32513	30887	29343	27876	26482	25158	23900	22705	21570	20492	19467	18494	17569	16691	
CLASS 29 CURRENT	0	0	0	0	0	0	0	1035836	2071672	1035836	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	
CLASS 29 CARRIED	0	0	0	0	0	0	0	1035836	2961680	3685324	3685324	3261401	2799058	2294853	1745030	1145495	491786	0	0	0	0	0	0	0	0	0	0	

Alternate Case 2
($80.00/tonne Coal)
September, 1981 Oil Pricing Agreement
Figure 3.7

```
                                                                                                                  ($1000)

                                  1982     1983     1984     1985     1986     1987     1988     1989     1990     1991     1992     1993     1994     1995     1996     1997     1998     1999     2000     2001     2002     2003     2004     2005     2006     2007     2008       TOTAL
                                  ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----       =====
************************
CASH FLOW CALCULATIONS
************************
+ REVENUE                            0        0        0        0        0        0        0   523036   997691  1398195  1693370  1845773  2011893  2192963  2390330  2605459  2839951  3095546  3374145  3677818  4008822  4369616  4762881  5191541  5658779  6168078  6723196    65529075
- OPERATING COSTS                    0        0        0        0        0        0        0   523036   832046  1047173  1217850  1327457  1446928  1577151  1719095  1873814  2042457  2226278  2426643  2645041  2883095  3142573  3425405  3733691  4069723  4435998  4835238    47430694
- FEDERAL INCOME TAX                 0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0   192082   312460   340648   371370   404853   441347   481123   524475   571727   623229   679364     4942679
- PROVINCIAL INCOME TAX              0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    48020    78115    85162    92842   101213   110337   120281   131119   142932   155807   169841     1235670
- INTEREST PAID FROM REVENUE         0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0           0
- DEBT REPAYMENTS                    0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0           0
- CAPITAL COSTS FROM REVENUE         0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0           0
- EQUITY CAPITAL COSTS               0        0   139720   479447   884880  1290047  1527226   209278        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     4530598
--------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
= CASH FLOW                          0        0  -139720  -479447  -884880 -1290047 -1527226  -209278   165644   351021   475519   518316   564965   615811   671233   731646   557392   478693   521692   568565   619661   675359   736073   802255   874397   953034  1038752     7389434
  ACCUMULATED                        0        0  -139720  -619167 -1504047 -2794094 -4321320 -4530598 -4364953 -4013933 -3538414 -3020098 -2455133 -1839321 -1168087  -436441   120951   599644  1121336  1689902  2309563  2984922  3720995  4523250  5397647  6350682  7389434

DCF RATE OF RETURN (%)            8.05
PAYOUT PERIOD (YEARS)             9.78
EQUITY RATIO (%)                100.00


*******************
COMPONENT INFLATION
*******************
PRICE         RATE                5.00    11.00    11.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00
              FACTOR             1.050    1.166    1.294    1.410    1.537    1.675    1.826    1.991    2.170    2.365    2.578    2.810    3.063    3.338    3.639    3.966    4.323    4.712    5.136    5.599    6.103    6.652    7.250    7.903    8.614    9.390   10.235

CAPITAL       RATE                5.00    11.00    11.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00
              FACTOR             1.050    1.166    1.294    1.410    1.537    1.675    1.826    1.991    2.170    2.365    2.578    2.810    3.063    3.338    3.639    3.966    4.323    4.712    5.136    5.599    6.103    6.652    7.250    7.903    8.614    9.390   10.235

OPERATING     RATE                5.00    11.00    11.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00
              FACTOR             1.050    1.166    1.294    1.410    1.537    1.675    1.826    1.991    2.170    2.365    2.578    2.810    3.063    3.338    3.639    3.966    4.323    4.712    5.136    5.599    6.103    6.652    7.250    7.903    8.614    9.390   10.235


********************
CAPITAL EXPENDITURES
********************
PROCESS FACILITIES                   0        0    90000   200000   320000   420000   505405       -2        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     1535403
BUILDINGS + SERVICES                 0        0        0        0        0    12000    12000        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0       24000
INDIRECTS                            0        0    12000    70000   130700   170000   162400        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      545100
CONTINGENCY                          0        0     6000    70000   125000   168000   156495        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      525495
WORKING CAPITAL                      0        0        0        0        0        0        0    73360        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0       73360
--------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
TOTAL CAPITAL COSTS                  0        0   108000   340000   575700   770000   836300    73360        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     2703360
INFLATION                            0        0    31720   139447   309180   520047   690926    72665        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     1763985
CAPITALIZED COSTS                    0        0        0        0        0        0        0    63253        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0       63253
--------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
TOTAL CAPITAL                        0        0   139720   479447   884880  1290047  1527226   209278        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     4530598
ACCUMULATED                          0        0   139720   619167  1504047  2794094  4321320  4530598  4530598  4530598  4530598  4530598  4530598  4530598  4530598  4530598  4530598  4530598  4530598  4530598  4530598  4530598  4530598  4530598  4530598  4530598  4530598


***************
OPERATING COSTS
***************
OPERATING COSTS                      0        0        0        0        0        0        0   294540   383490   442790   472440   472440   472440   472440   472440   472440   472440   472440   472440   472440   472440   472440   472440   472440   472440   472440   472440     9152300
INFLATION                            0        0        0        0        0        0        0   291749   448556   604383   745410   855017   974488  1104711  1246655  1401374  1570017  1753838  1954203  2172601  2410655  2670133  2952965  3261251  3597283  3963558  4362798    38341647
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TOTAL OPERATING COSTS                0        0        0        0        0        0        0   586289   832046  1047173  1217850  1327457  1446928  1577151  1719095  1873814  2042457  2226278  2426643  2645041  2883095  3142573  3425405  3733691  4069723  4435998  4835238    47493947
CAPITALIZED COSTS                    0        0        0        0        0        0        0    63253        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0       63253
--------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
TOTAL OPERATING COSTS                0        0        0        0        0        0        0   523036   832046  1047173  1217850  1327457  1446928  1577151  1719095  1873814  2042457  2226278  2426643  2645041  2883095  3142573  3425405  3733691  4069723  4435998  4835238    47430694


******************
 TAX CALCULATIONS
******************

FEDERAL INCOME TAX
******************
+ REVENUE                            0        0        0        0        0        0        0   523036   997691  1398195  1693370  1845773  2011893  2192963  2390330  2605459  2839951  3095546  3374145  3677818  4008822  4369616  4762881  5191541  5658779  6168078  6723196    65529075
- OPERATING COSTS                    0        0        0        0        0        0        0   523036   832046  1047173  1217850  1327457  1446928  1577151  1719095  1873814  2042457  2226278  2426643  2645041  2883095  3142573  3425405  3733691  4069723  4435998  4835238    47430694
- INTEREST                           0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0           0
- CLASS 3                            0        0        0        0        0        0        0        0     2101     1996     1896     1801     1711     1626     1544     1467     1394     1324     1258     1195     1135     1078     1025      973      925      878      835       26163
- CLASS 29 CURRENT                   0        0        0        0        0        0        0        0   163543   349026        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      512568
- CLASS 29 CARRIED                   0        0        0        0        0        0        0        0        0        0   473623   516515   563253   614186   669690   730178   262540        0        0        0        0        0        0        0        0        0        0     3829986
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= TAXABLE INCOME                     0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0   533560   867944   946244  1031582  1124592  1225964  1336452  1456876  1588131  1731193  1887123    13729663
  INCOME TAX (36.0%)                 0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0   192082   312460   340648   371370   404853   441347   481123   524475   571727   623229   679364     4942679


PROVINCIAL INCOME TAX
*********************
FEDERAL TAXABLE INCOME               0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0   533560   867944   946244  1031582  1124592  1225964  1336452  1456876  1588131  1731193  1887123    13729663
INCOME TAX (9.0 %)                   0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    48020    78115    85162    92842   101213   110337   120281   131119   142932   155807   169841     1235670


CAPITAL COST ALLOWANCE
**********************
INPUT
   CLASS 3                           0        0        0        0        0    20105    21914        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0       42019
   CLASS 29                          0        0   139720   479447   884880  1269942  1505312    63253        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     4342555
   NO ALLOWANCE                      0        0        0        0        0        0        0   146025        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      146025
--------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
   TOTAL INPUT                       0        0   139720   479447   884880  1290047  1527226   209278        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     4530598

POOLS BEFORE DEDUCTIONS
   CLASS 3                           0        0        0        0        0    20105    42019    42019    42019    39918    37922    36026    34224    32513    30888    29343    27876    26482    25158    23900    22705    21570    20491    19467    18494    17569    16691
   CLASS 29 CURRENT                  0        0        0        0        0        0        0  1085639  2171277  1085639        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0
   CLASS 29 CARRIED                  0        0        0        0        0        0        0  1085639  3093373  3829986  3829986  3356363  2839848  2276595  1662409   992719   262540        0        0        0        0        0        0        0        0        0        0
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Alternate Case 3
($80.00/tonne Coal)
September, 1981 Oil Pricing Agreement
Figure 3.8

($1000)

CASH FLOW CALCULATIONS

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
+ REVENUE	0	0	0	0	0	0	0	625521	1193181	1672159	2025170	2207435	2406184	2622654	2858692	3115975	3396412	3702090	4035278	4398453	4794313	5225801	5696124	6208775	6767554	7376645	8040543	78360889
- OPERATING COSTS	0	0	0	0	0	0	0	625521	995164	1283337	1507078	1642715	1790560	1951710	2127364	2318827	2527521	2754998	3002948	3273213	3567802	3888904	4238906	4620407	5036244	5489506	5983561	58626282
- FEDERAL INCOME TAX	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	150431	340476	371186	404656	441135	480895	524230	571462	622943	679054	740213	5326680
- PROVINCIAL INCOME TAX	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	37608	85119	92796	101164	110284	120224	131057	142865	155736	169764	185053	1331670
- INTEREST PAID FROM REVENUE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
- DEBT REPAYMENTS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
- CAPITAL COSTS FROM REVENUE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
- EQUITY CAPITAL COSTS	0	0	194961	637524	1158168	1534988	1399759	197761	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	5123161
= CASH FLOW	0	0	-194961	-637524	-1158168	-1534988	-1399759	-197761	198017	388823	518092	564720	615545	670944	731329	797148	680853	521496	568348	619420	675092	735779	801931	874040	952643	1038322	1131716	7961095
ACCUMULATED	0	0	-194961	-832485	-1990653	-3525641	-4925400	-5123161	-4925144	-4536321	-4018229	-3453510	-2837965	-2167021	-1435693	-638545	42309	563805	1132153	1751572	2426664	3162443	3964374	4838414	5791057	6829379	7961095	

DCF RATE OF RETURN (%) 7.70
PAYOUT PERIOD (YEARS) 9.94
EQUITY RATIO (%) 100.00

COMPONENT INFLATION

		1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008
PRICE	RATE	5.00	11.00	11.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00
	FACTOR	1.050	1.166	1.294	1.410	1.537	1.675	1.826	1.991	2.170	2.365	2.578	2.810	3.063	3.338	3.639	3.966	4.323	4.712	5.136	5.599	6.103	6.652	7.250	7.903	8.614	9.390	10.235
CAPITAL	RATE	5.00	11.00	11.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00
	FACTOR	1.050	1.166	1.294	1.410	1.537	1.675	1.826	1.991	2.170	2.365	2.578	2.810	3.063	3.338	3.639	3.966	4.323	4.712	5.136	5.599	6.103	6.652	7.250	7.903	8.614	9.390	10.235
OPERATING	RATE	5.00	11.00	11.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00
	FACTOR	1.050	1.166	1.294	1.410	1.537	1.675	1.826	1.991	2.170	2.365	2.578	2.810	3.063	3.338	3.639	3.966	4.323	4.712	5.136	5.599	6.103	6.652	7.250	7.903	8.614	9.390	10.235

CAPITAL EXPENDITURES

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
PROCESS FACILITIES	0	0	88800	266400	444000	532800	444810	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1776810
BUILDINGS + SERVICES	0	0	0	0	0	12000	12000	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	24000
INDIRECTS	0	0	31500	94500	157500	189000	157690	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	630190
CONTINGENCY	0	0	30400	91200	152000	182400	152000	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	608000
WORKING CAPITAL	0	0	0	0	0	0	0	80900	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	80900
TOTAL CAPITAL COSTS	0	0	150700	452100	753500	916200	766500	80900	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	3119900
INFLATION	0	0	44261	185424	404668	618788	633259	80134	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1966534
CAPITALIZED COSTS	0	0	0	0	0	0	0	36727	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	36727
TOTAL CAPITAL	0	0	194961	637524	1158168	1534988	1399759	197761	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	5123161
ACCUMULATED	0	0	194961	832485	1990653	3525641	4925400	5123161	5123161	5123161	5123161	5123161	5123161	5123161	5123161	5123161	5123161	5123161	5123161	5123161	5123161	5123161	5123161	5123161	5123161	5123161	5123161	

OPERATING COSTS

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
OPERATING COSTS	0	0	0	0	0	0	0	332780	458670	542650	584640	584640	584640	584640	584640	584640	584640	584640	584640	584640	584640	584640	584640	584640	584640	584640	584640	11272980
INFLATION	0	0	0	0	0	0	0	329468	536494	740687	922438	1058075	1205920	1367070	1542724	1734187	1942881	2170358	2418308	2688573	2983162	3304264	3654266	4035767	4451604	4904866	5398921	47390109
TOTAL OPERATING COSTS	0	0	0	0	0	0	0	662248	995164	1283337	1507078	1642715	1790560	1951710	2127364	2318827	2527521	2754998	3002948	3273213	3567802	3888904	4238906	4620407	5036244	5489506	5983561	58663089
CAPITALIZED COSTS	0	0	0	0	0	0	0	36727	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	36727
TOTAL OPERATING COSTS	0	0	0	0	0	0	0	625521	995164	1283337	1507078	1642715	1790560	1951710	2127364	2318827	2527521	2754998	3002948	3273213	3567802	3888904	4238906	4620407	5036244	5489506	5983561	58626282

TAX CALCULATIONS

FEDERAL INCOME TAX

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
+ REVENUE	0	0	0	0	0	0	0	625521	1193181	1672159	2025170	2207435	2406184	2622654	2858692	3115975	3396412	3702090	4035278	4398453	4794313	5225801	5696124	6208775	6767554	7376645	8040543	78360889
- OPERATING COSTS	0	0	0	0	0	0	0	625521	995164	1283337	1507078	1642715	1790560	1951710	2127364	2318827	2527521	2754998	3002948	3273213	3567802	3888904	4238906	4620407	5036244	5489506	5983561	58626282
- INTEREST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
- CLASS 3	0	0	0	0	0	0	0	0	2101	1996	1896	1801	1711	1626	1544	1467	1394	1324	1258	1195	1135	1078	1025	973	925	878	835	26163
- CLASS 29 CURRENT	0	0	0	0	0	0	0	0	195916	386827	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	582743
- CLASS 29 CARRIED	0	0	0	0	0	0	0	0	0	0	516196	562919	613833	669318	729754	795681	449635	0	0	0	0	0	0	0	0	0	0	4337366
= TAXABLE INCOME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	417863	945768	1031072	1124045	1225376	1335819	1456193	1587394	1730396	1886261	2056148	14796335
INCOME TAX (36.0%)	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	150431	340476	371186	404656	441135	480895	524230	571462	622943	679054	740213	5326680

PROVINCIAL INCOME TAX

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
FEDERAL TAXABLE INCOME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	417863	945768	1031072	1124045	1225376	1335819	1456193	1587394	1730396	1886261	2056148	14796335
INCOME TAX (9.0 %)	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	37608	85119	92796	101164	110284	120224	131057	142865	155736	169764	185053	1331670

CAPITAL COST ALLOWANCE

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
INPUT																												
CLASS 3	0	0	0	0	0	20105	21914	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	42019
CLASS 29	0	0	194961	637524	1158168	1514884	1377845	36727	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	4920109
NO ALLOWANCE	0	0	0	0	0	0	0	161034	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	161034
TOTAL INPUT	0	0	194961	637524	1158168	1534988	1399759	197761	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	5123161
POOLS BEFORE DEDUCTIONS																												
CLASS 3	0	0	0	0	0	20105	42019	42019	42019	39918	37922	36026	34224	32513	30888	29343	27876	26482	25158	23900	22705	21570	20491	19467	18494	17569	16690	
CLASS 29 CURRENT	0	0	0	0	0	0	0	1230827	2460854	1230827	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	
CLASS 29 CARRIED	0	0	0	0	0	0	0	1230827	3494165	4337366	4337366	3821170	3258252	2644418	1975100	1245316	449635	0	0	0	0	0	0	0	0	0	0	

Alternate Case 4
($80.00/tonne Coal)
September, 1981 Oil Pricing Agreement
Figure 3.9

```
                                                                                                                        ($1000)

                                     1982     1983     1984     1985     1986     1987     1988     1989     1990     1991     1992     1993     1994     1995     1996     1997     1998     1999     2000     2001     2002     2003     2004     2005     2006     2007     2008      TOTAL
                                     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----      =====
**********************
CASH FLOW CALCULATIONS
**********************
+ REVENUE                               0        0        0        0        0        0        0   382213   729071  1021741  1237441  1348811  1470284  1602523  1746730  1903957  2075313  2262091  2465580  2687591  2929474  3193127  3480508  3793754  4135192  4507359  4913021   47885820
- OPERATING COSTS                       0        0        0        0        0        0        0   382213   663117   860674  1013277  1104472  1203874  1312223  1430323  1559052  1699367  1852310  2019018  2200730  2398795  2614687  2850009  3106509  3386095  3690844  4023020   39370608
- FEDERAL INCOME TAX                    0        0   -20083   -87545  -201766  -320628  -372911  -243447   -61152    53134    80233    87519    95458   104108   113534   123805   134998   147196   160489   174976   190765   207973   226728   247169   269447   293730   320195    1723929
- PROVINCIAL INCOME TAX                 0        0    -2734   -14409   -36858   -62389   -77089   -59950   -15288    13284    20058    21880    23865    26027    28383    30951    33750    36799    40122    43744    47691    51993    56682    61792    67362    73432    80049     489147
- INTEREST PAID FROM REVENUE            0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0          0
- DEBT REPAYMENTS                       0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0          0
- CAPITAL COSTS FROM REVENUE            0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0          0
- EQUITY CAPITAL COSTS                  0        0   130664   427272   776211  1015284   922216   162534        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    3434181
-------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
= CASH FLOW                             0        0  -107847  -325318  -537588  -632267  -472216   140863   142394    94649   123873   134940   147007   160164   174509   190148   207199   225787   246050   268141   292222   318473   347089   378283   412257   449353   489787    2867955
  ACCUMULATED                           0        0  -107847  -433165  -970753 -1603019 -2075235 -1934372 -1791978 -1697329 -1573456 -1438516 -1291509 -1131345  -956836  -766688  -559489  -333703   -87652   180489   472711   791184  1138274  1516557  1928814  2378167  2867955

DCF RATE OF RETURN (%)               6.63
PAYOUT PERIOD (YEARS)               12.33
EQUITY RATIO (%)                   100.00


*******************
COMPONENT INFLATION
*******************
PRICE        RATE                    5.00    11.00    11.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00
             FACTOR                 1.050    1.166    1.294    1.410    1.537    1.675    1.826    1.991    2.170    2.365    2.578    2.810    3.063    3.338    3.639    3.966    4.323    4.712    5.136    5.599    6.103    6.652    7.250    7.903    8.614    9.390   10.235

CAPITAL      RATE                    5.00    11.00    11.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00
             FACTOR                 1.050    1.166    1.294    1.410    1.537    1.675    1.826    1.991    2.170    2.365    2.578    2.810    3.063    3.338    3.639    3.966    4.323    4.712    5.136    5.599    6.103    6.652    7.250    7.903    8.614    9.390   10.235

OPERATING    RATE                    5.00    11.00    11.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00
             FACTOR                 1.050    1.166    1.294    1.410    1.537    1.675    1.826    1.991    2.170    2.365    2.578    2.810    3.063    3.338    3.639    3.966    4.323    4.712    5.136    5.599    6.103    6.652    7.250    7.903    8.614    9.390   10.235


********************
CAPITAL EXPENDITURES
********************
PROCESS FACILITIES                      0        0    86000   180000   300000   350000   264000        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    1180000
BUILDINGS + SERVICES                    0        0        0        0        0    10000    10000        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      20000
INDIRECTS                               0        0    10000    63000   105000   126000   116000        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     420000
CONTINGENCY                             0        0     5000    60000   100000   120000   115000        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     400000
WORKING CAPITAL                         0        0        0        0        0        0        0    55490        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      55490
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TOTAL CAPITAL COSTS                     0        0   101000   303000   505000   606000   505000    55490        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    2075490
INFLATION                               0        0    29664   124272   271211   409284   417216    54964        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    1306611
CAPITALIZED COSTS                       0        0        0        0        0        0        0    52080        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      52080
-------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
TOTAL CAPITAL                           0        0   130664   427272   776211  1015284   922216   162534        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    3434181
ACCUMULATED                             0        0   130664   557936  1334147  2349430  3271646  3434181  3434181  3434181  3434181  3434181  3434181  3434181  3434181  3434181  3434181  3434181  3434181  3434181  3434181  3434181  3434181  3434181  3434181  3434181  3434181


***************
OPERATING COSTS
***************
OPERATING COSTS                         0        0        0        0        0        0        0   218180   305630   363930   393080   393080   393080   393080   393080   393080   393080   393080   393080   393080   393080   393080   393080   393080   393080   393080   393080    7570100
INFLATION                               0        0        0        0        0        0        0   216113   357487   496744   620197   711392   810794   919143  1037243  1165972  1306287  1459230  1625938  1807650  2005715  2221607  2456929  2713429  2993015  3297764  3629940   31857580
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TOTAL OPERATING COSTS                   0        0        0        0        0        0        0   434293   663117   860674  1013277  1104472  1203874  1312223  1430323  1559052  1699367  1852310  2019018  2200730  2398795  2614687  2850009  3106509  3386095  3690844  4023020   39422688
CAPITALIZED COSTS                       0        0        0        0        0        0        0    52080        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      52080
-------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
TOTAL OPERATING COSTS                   0        0        0        0        0        0        0   382213   663117   860674  1013277  1104472  1203874  1312223  1430323  1559052  1699367  1852310  2019018  2200730  2398795  2614687  2850009  3106509  3386095  3690844  4023020   39370608


******************
FEDERAL INCOME TAX
******************
+ REVENUE                               0        0        0        0        0        0        0   382213   729071  1021741  1237441  1348811  1470284  1602523  1746730  1903957  2075313  2262091  2465580  2687591  2929474  3193127  3480508  3793754  4135192  4507359  4913021   47885820
- OPERATING COSTS                       0        0        0        0        0        0        0   382213   663117   860674  1013277  1104472  1203874  1312223  1430323  1559052  1699367  1852310  2019018  2200730  2398795  2614687  2850009  3106509  3386095  3690844  4023020   39370608
- INTEREST                              0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0          0
- CLASS 3                               0        0        0        0        0      779     1589     1510     1434     1363     1294     1230     1168     1110     1054     1002      952      904      859      816      775      736      699      665      631      600      570      21740
- CLASS 29 CURRENT                      0        0    30379   160100   409530   692437   854955   664606   234387    12109        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    3058501
- CLASS 29 CARRIED                      0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0          0
-------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
= TAXABLE INCOME                        0        0   -30379  -160100  -409530  -693216  -856544  -666115  -169867   147596   222870   243110   265162   289190   315372   343903   374995   408877   445803   486045   529904   577704   629800   686580   748453   815915   889432    5434971
  INCOME TAX (36.0%)                    0        0   -10937   -57636  -147431  -249558  -308356  -239802   -61152    53134    80233    87519    95458   104108   113534   123805   134998   147196   160489   174976   190765   207973   226728   247169   269447   293730   320195    1956589
- INVESTMENT TAX CREDIT                 0        0     9146    29909    54335    71070    64555     3646        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     232661
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= TAX PAYABLE AFTER ITC                 0        0   -20083   -87545  -201766  -320628  -372911  -243447   -61152    53134    80233    87519    95458   104108   113534   123805   134998   147196   160489   174976   190765   207973   226728   247169   269447   293730   320195    1723929


*********************
PROVINCIAL INCOME TAX
*********************
FEDERAL TAXABLE INCOME                  0        0   -30379  -160100  -409530  -693216  -856544  -666115  -169867   147596   222870   243110   265162   289190   315372   343903   374995   408877   445803   486045   529904   577704   629800   686580   748453   815915   889432    5434971
INCOME TAX (9.0 %)                      0        0    -2734   -14409   -36858   -62389   -77089   -59950   -15288    13284    20058    21880    23865    26027    28383    30951    33750    36799    40122    43744    47691    51993    56682    61792    67362    73432    80049     489147

CAPITAL COST ALLOWANCE
**********************
INPUT
   CLASS 3  - CCA                       0        0        0        0        0    15581    16983        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      32564
             ITC                        0        0        0        0        0     1173     1278        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0       2451
   CLASS 29 - CCA                       0        0   121518   397363   721876   928633   848677    48435        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    3058501
             ITC                        0        0     9146    29909    54335    69897    63277     3646        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     230210
   NO ALLOWANCE                         0        0        0        0        0        0        0   110454        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     110454
-------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
   TOTAL INPUT                          0        0   130664   427272   776211  1015284   922216   162534        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    3434181

POOLS BEFORE DEDUCTIONS
   CLASS 3                              0        0        0        0        0    15581    31705    30196    28686    27252    25889    24595    23365    22197    21087    20033    19031    18080    17176    16317    15501    14726    13990    13290    12626    11994    11395
   CLASS 29 CURRENT                     0        0    30379   160100   409530   692437   854955   664606   234387    12109        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0
   CLASS 29 CARRIED                     0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0

INVESTMENT TAX CREDIT (7.0
**************************
INPUT: CLASS 3                          0        0        0        0        0     1173     1278        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0       2451
       CLASS 29                         0        0     9146    29909    54335    69897    63277     3646        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     230210
-------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
       TOTAL                            0        0     9146    29909    54335    71070    64555     3646        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     232661

POOL BEFORE DEDUCTION                   0        0     9146    29909    54335    71070    64555     3646        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     232661
```

Special Case Corporate Tax Calculations
Figure 3.10

($1000)

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
CASH FLOW CALCULATIONS																												
+ REVENUE	0	0	0	0	0	0	0	356191	673204	934790	1121749	1211489	1308408	1413081	1526129	1648218	1780075	1922481	2076280	2242382	2421773	2615515	2824756	3050736	3294795	3558379	3843049	39823480
- OPERATING COSTS	0	0	0	0	0	0	0	345451	488601	614313	706273	755712	808612	865214	925780	990584	1059925	1134120	1213508	1298454	1389345	1486600	1590662	1702008	1821149	1948629	2085033	23229970
- FEDERAL INCOME TAX	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	159531	258857	283433	310239	339474	371350	406102	443982	485265	530249	579260	632648	4800390
- PROVINCIAL INCOME TAX	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	39883	64714	70858	77560	84868	92838	101526	110995	121316	132562	144815	158162	1200097
- INTEREST PAID FROM REVENUE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
- DEBT REPAYMENTS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
- CAPITAL COSTS FROM REVENUE	0	0	0	0	0	0	0	10741	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	10741
- EQUITY CAPITAL COSTS	0	0	130664	427272	776211	1015284	922216	96847	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	3368494
= CASH FLOW	0	0	-130664	-427272	-776211	-1015284	-922216	-96847	184603	320477	415477	455777	499797	547867	600347	458220	396579	434071	474973	519586	568239	621287	679117	742147	810835	885675	967206	7213788
ACCUMULATED	0	0	-130664	-557936	-1334147	-2349430	-3271646	-3368494	-3183891	-2863414	-2447937	-1992160	-1492363	-944496	-344149	114072	510651	944721	1419694	1939280	2507520	3128807	3807924	4550072	5360907	6246582	7213788	
DCF RATE OF RETURN (%)	9.52																											
PAYOUT PERIOD (YEARS)	8.75																											
EQUITY RATIO (%)	100.00																											
COMPONENT INFLATION																												
PRICE RATE	5.00	10.00	10.00	8.00	8.00	8.00	8.00	8.00	8.00	8.00	8.00	8.00	8.00	8.00	8.00	8.00	8.00	8.00	8.00	8.00	8.00	8.00	8.00	8.00	8.00	8.00	8.00	
PRICE FACTOR	1.050	1.155	1.271	1.372	1.482	1.600	1.729	1.867	2.016	2.177	2.352	2.540	2.743	2.962	3.199	3.455	3.732	4.030	4.353	4.701	5.077	5.483	5.922	6.395	6.907	7.460	8.056	
CAPITAL RATE	5.00	11.00	11.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	
CAPITAL FACTOR	1.050	1.166	1.294	1.410	1.537	1.675	1.826	1.991	2.170	2.365	2.578	2.810	3.063	3.338	3.639	3.966	4.323	4.712	5.136	5.599	6.103	6.652	7.250	7.903	8.614	9.390	10.235	
OPERATING RATE	4.50	9.00	9.00	7.00	7.00	7.00	7.00	7.00	7.00	7.00	7.00	7.00	7.00	7.00	7.00	7.00	7.00	7.00	7.00	7.00	7.00	7.00	7.00	7.00	7.00	7.00	7.00	
OPERATING FACTOR	1.045	1.139	1.242	1.328	1.421	1.521	1.627	1.741	1.863	1.994	2.133	2.283	2.442	2.613	2.795	2.992	3.201	3.426	3.665	3.922	4.196	4.490	4.804	5.141	5.501	5.886	6.298	
CAPITAL EXPENDITURES																												
PROCESS FACILITIES	0	0	86000	180000	300000	350000	264000	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1180000
BUILDINGS + SERVICES	0	0	0	0	0	10000	10000	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	20000
INDIRECTS	0	0	10000	63000	105000	126000	116000	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	420000
CONTINGENCY	0	0	5000	60000	100000	120000	115000	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	400000
WORKING CAPITAL	0	0	0	0	0	0	0	54050	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	54050
TOTAL CAPITAL COSTS	0	0	101000	303000	505000	606000	505000	54050	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	2074050
INFLATION	0	0	29664	124272	271211	409284	417216	53538	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1305184
CAPITALIZED COSTS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
TOTAL CAPITAL	0	0	130664	427272	776211	1015284	922216	107588	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	3379234
ACCUMULATED	0	0	130664	557936	1334147	2349430	3271646	3379234	3379234	3379234	3379234	3379234	3379234	3379234	3379234	3379234	3379234	3379234	3379234	3379234	3379234	3379234	3379234	3379234	3379234	3379234	3379234	
OPERATING COSTS																												
OPERATING COSTS	0	0	0	0	0	0	0	198380	262230	308130	331080	331080	331080	331080	331080	331080	331080	331080	331080	331080	331080	331080	331080	331080	331080	331080	331080	6397100
INFLATION	0	0	0	0	0	0	0	147071	226371	306183	375193	424632	477532	534134	594700	659504	728845	803040	882428	967374	1058265	1155520	1259582	1370928	1490069	1617549	1753953	16832870
TOTAL OPERATING COSTS	0	0	0	0	0	0	0	345451	488601	614313	706273	755712	808612	865214	925780	990584	1059925	1134120	1213508	1298454	1389345	1486600	1590662	1702008	1821149	1948629	2085033	23229970
CAPITALIZED COSTS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
TOTAL OPERATING COSTS	0	0	0	0	0	0	0	345451	488601	614313	706273	755712	808612	865214	925780	990584	1059925	1134120	1213508	1298454	1389345	1486600	1590662	1702008	1821149	1948629	2085033	23229970
FEDERAL INCOME TAX																												
+ REVENUE	0	0	0	0	0	0	0	356191	673204	934790	1121749	1211489	1308408	1413081	1526129	1648218	1780075	1922481	2076280	2242382	2421773	2615515	2824756	3050736	3294795	3558379	3843049	39823480
- OPERATING COSTS	0	0	0	0	0	0	0	345451	488601	614313	706273	755712	808612	865214	925780	990584	1059925	1134120	1213508	1298454	1389345	1486600	1590662	1702008	1821149	1948629	2085033	23229970
- INTEREST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
- CLASS 3	0	0	0	0	0	0	0	1751	1663	1580	1501	1426	1355	1287	1223	1162	1103	1048	996	946	899	854	811	771	732	695	661	22463
- CLASS 29 CURRENT	0	0	0	0	0	0	0	8990	182940	318897	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	510826
- CLASS 29 CARRIED	0	0	0	0	0	0	0	0	0	0	413976	454351	498442	546580	599126	213330	0	0	0	0	0	0	0	0	0	0	0	2725805
= TAXABLE INCOME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	443142	719047	787313	861776	942982	1031529	1128061	1233283	1347958	1472915	1609054	1757356	13334416
INCOME TAX (36.0%)	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	159531	258857	283433	310239	339474	371350	406102	443982	485265	530249	579260	632648	4800390
- INVESTMENT TAX CREDIT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
= TAX PAYABLE AFTER ITC	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	159531	258857	283433	310239	339474	371350	406102	443982	485265	530249	579260	632648	4800390
PROVINCIAL INCOME TAX																												
FEDERAL TAXABLE INCOME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	443142	719047	787313	861776	942982	1031529	1128061	1233283	1347958	1472915	1609054	1757356	13334416
INCOME TAX (9.0 %)	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	39883	64714	70858	77560	84868	92838	101526	110995	121316	132562	144815	158162	1200097
CAPITAL COST ALLOWANCE																												
INPUT																												
CLASS 3 - CCA	0	0	0	0	0	16754	18262	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	35016
ITC	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CLASS 29 - CCA	0	0	130664	427272	776211	998530	903954	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	3236631
ITC	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO ALLOWANCE	0	0	0	0	0	0	0	107588	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	107588
TOTAL INPUT	0	0	130664	427272	776211	1015284	922216	107588	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	3379234
POOLS BEFORE DEDUCTIONS																												
CLASS 3	0	0	0	0	0	16754	35016	35016	33265	31602	30021	28520	27094	25740	24453	23230	22069	20965	19917	18921	17975	17076	16222	15411	14641	13909	13213	
CLASS 29 CURRENT	0	0	0	0	0	0	0	809158	1618315	809158	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	
CLASS 29 CARRIED	0	0	0	0	0	0	0	800168	2235544	2725805	2725805	2311829	1857478	1359036	812456	213330	0	0	0	0	0	0	0	0	0	0	0	

Special Case
Syncrude Discount
($5.00/bbl,
$60.00/tonne Coal)
Modified Escalation
Figure 3.11

```
                                                                                                                                   ($1000)

                                          1982      1983      1984      1985      1986      1987      1988      1989      1990      1991      1992      1993      1994      1995      1996      1997      1998      1999      2000      2001      2002      2003      2004      2005      2006      2007      2008       TOTAL
                                          ----      ----      ----      ----      ----      ----      ----      ----      ----      ----      ----      ----      ----      ----      ----      ----      ----      ----      ----      ----      ----      ----      ----      ----      ----      ----      ----       =====

*********************
CASH FLOW CALCULATIONS
*********************
+ REVENUE                                    0         0         0         0         0         0         0    342124    652601    914573   1107650   1207338   1315999   1434439   1563538   1704257   1857640   2024827   2207062   2405697   2622210   2858209   3115448   3395838   3701464   4034595   4397709    42863215
- OPERATING COSTS                            0         0         0         0         0         0         0    342124    652601    857398   1009784   1100665   1199724   1307700   1425393   1553678   1693509   1845925   2012058   2193143   2390526   2605674   2840184   3095801   3374423   3678121   4009152    39187582
- FEDERAL INCOME TAX                         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0     90725       90725
- PROVINCIAL INCOME TAX                      0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0     22681       22681
- INTEREST PAID FROM REVENUE                 0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0           0
- DEBT REPAYMENTS                            0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0           0
- CAPITAL COSTS FROM REVENUE                 0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0           0
- EQUITY CAPITAL COSTS                       0         0    130664    437143    791581   1032037    949608    210815      7381         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0     3559230
-------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
= CASH FLOW                                  0         0   -130664   -437143   -791581  -1032037   -949608   -210815     -7381     57175     97866    106674    116274    126739    138145    150579    164131    178902    195004    212554    231684    252535    275263    300037    327041    356474    275151        2998
  ACCUMULATED                                0         0   -130664   -567807  -1359388  -2391426  -3341034  -3551849  -3559230  -3502055  -3404189  -3297516  -3181242  -3054503  -2916357  -2765779  -2601648  -2422746  -2227742  -2015188  -1783504  -1530969  -1255705   -955668   -628628   -272153      2998

DCF RATE OF RETURN (%)                     .01
PAYOUT PERIOD (YEARS)                    19.99
EQUITY RATIO (%)                        100.00


*******************
COMPONENT INFLATION
*******************
PRICE        RATE                         5.00     11.00     11.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00
             FACTOR                      1.050     1.166     1.294     1.410     1.537     1.675     1.826     1.991     2.170     2.365     2.578     2.810     3.063     3.338     3.639     3.966     4.323     4.712     5.136     5.599     6.103     6.652     7.250     7.903     8.614     9.390    10.235

CAPITAL      RATE                         5.00     11.00     11.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00
             FACTOR                      1.050     1.166     1.294     1.410     1.537     1.675     1.826     1.991     2.170     2.365     2.578     2.810     3.063     3.338     3.639     3.966     4.323     4.712     5.136     5.599     6.103     6.652     7.250     7.903     8.614     9.390    10.235

OPERATING    RATE                         5.00     11.00     11.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00      9.00
             FACTOR                      1.050     1.166     1.294     1.410     1.537     1.675     1.826     1.991     2.170     2.365     2.578     2.810     3.063     3.338     3.639     3.966     4.323     4.712     5.136     5.599     6.103     6.652     7.250     7.903     8.614     9.390    10.235


********************
CAPITAL EXPENDITURES
********************
PROCESS FACILITIES                           0         0     86000    185000    305000    355000    271160         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0     1202160
BUILDINGS + SERVICES                         0         0         0         0         0     10000     10000         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0       20000
INDIRECTS                                    0         0     10000     63000    107000    128000    119740         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0      427740
CONTINGENCY                                  0         0      5000     62000    103000    123000    119100         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0      412100
WORKING CAPITAL                              0         0         0         0         0         0         0     61140         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0       61140
-------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
TOTAL CAPITAL COSTS                          0         0    101000    310000    515000    616000    520000     61140         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0     2123140
INFLATION                                    0         0     29664    127143    276581    416037    429608     60561         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0     1339595
CAPITALIZED COSTS                            0         0         0         0         0         0         0     89114      7381         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0       96495
-------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
TOTAL CAPITAL                                0         0    130664    437143    791581   1032037    949608    210815      7381         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0     3559230
ACCUMULATED                                  0         0    130664    567807   1359388   2391426   3341034   3551849   3559230   3559230   3559230   3559230   3559230   3559230   3559230   3559230   3559230   3559230   3559230   3559230   3559230   3559230   3559230   3559230   3559233   3559230   3559230


***************
OPERATING COSTS
***************
OPERATING COSTS                              0         0         0         0         0         0         0    216645    304185    362545    391725    391725    391725    391725    391725    391725    391725    391725    391725    391725    391725    391725    391725    391725    391725    391725    391725     7542700
INFLATION                                    0         0         0         0         0         0         0    214592    355797    494853    618059    708940    807999    915975   1033668   1161953   1301784   1454200   1620333   1801418   1998801   2213949   2448459   2704076   2982698   3286396   3617427    31741377
-------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
TOTAL OPERATING COSTS                        0         0         0         0         0         0         0    431237    659982    857398   1009784   1100665   1199724   1307700   1425393   1553678   1693509   1845925   2012058   2193143   2390526   2605674   2840184   3095801   3374423   3678121   4009152    39284077
CAPITALIZED COSTS                            0         0         0         0         0         0         0     89114      7381         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0       96495
-------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
TOTAL OPERATING COSTS                        0         0         0         0         0         0         0    342124    652601    857398   1009784   1100665   1199724   1307700   1425393   1553678   1693509   1845925   2012058   2193143   2390526   2605674   2840184   3095801   3374423   3678121   4009152    39187582


******************
FEDERAL INCOME TAX
******************
+ REVENUE                                    0         0         0         0         0         0         0    342124    652601    914573   1107650   1207338   1315999   1434439   1563538   1704257   1857640   2024827   2207062   2405697   2622210   2858209   3115448   3395838   3701464   4034595   4397709    42863215
- OPERATING COSTS                            0         0         0         0         0         0         0    342124    652601    857398   1009784   1100665   1199724   1307700   1425393   1553678   1693509   1845925   2012058   2193143   2390526   2605674   2840184   3095801   3374423   3678121   4009152    39187582
- INTEREST                                   0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0           0
- CLASS 3                                    0         0         0         0         0         0         0         0         0      1751      1663      1580      1501      1426      1354      1287      1223      1162      1103      1048       996       946       899       854       811       771       732       21107
- CLASS 29 CURRENT                           0         0         0         0         0         0         0         0         0     55424      1845         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0       57269
- CLASS 29 CARRIED                           0         0         0         0         0         0         0         0         0         0     94357    105094    114773    125313    136791    149292    162908    177741    193900    211506    230688    251589    274365    299183    326229    355704    135812     3345244
-------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
= TAXABLE INCOME                             0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0    252013      252013
  INCOME TAX (36.0%)                         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0     90725       90725
- INVESTMENT TAX CREDIT                      0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0           0
-------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
= TAX PAYABLE AFTER ITC                      0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0     90725       90725


*********************
PROVINCIAL INCOME TAX
*********************
FEDERAL TAXABLE INCOME                       0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0    252013      252013
INCOME TAX (9.0 %)                           0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0     22681       22681


CAPITAL COST ALLOWANCE
***********************
INPUT
   CLASS 3  - CCA                            0         0         0         0         0     16754     18262         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0       35016
              ITC                            0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0           0
   CLASS 29 - CCA                            0         0    130664    437143    791581   1015284    931347     89114      7381         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0     3402514
              ITC                            0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0           0
   NO ALLOWANCE                              0         0         0         0         0         0         0    121701         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0      121701
-------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
   TOTAL INPUT                               0         0    130664    437143    791581   1032037    949608    210815      7381         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0     3559230

POOLS BEFORE DEDUCTIONS
   CLASS 3                                   0         0         0         0         0     16754     35016     35016     35016     35016     33265     31602     30021     28520     27094     25740     24453     23230     22069     20965     19917     18921     17975     17076     16222     15411     14641
   CLASS 29 CURRENT                          0         0         0         0         0         0         0    840783   1699412    852470      1845         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0         0
   CLASS 29 CARRIED                          0         0         0         0         0         0         0    840783   2540195   3345244   3345244   3250887   3145794   3031020   2905708   2768917   2619625   2456717   2278976   2085076   1873570   1642883   1391293   1116929    817745    491516    135812
```

Alternate Case 1
($80.00/tonne Coal)
April 13, 1982
Announcement
Figure 3.12

```
                                   1982     1983     1984     1985     1986     1987     1988     1989     1990     1991     1992     1993     1994     1995     1996     1997     1998     1999     2000     2001     2002     2003     2004     2005     2006     2007     2008     TOTAL
                                   ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     *****
**********************
CASH FLOW CALCULATIONS
**********************
+ REVENUE                             0        0        0        0        0        0        0   305309   582375   816156   988458  1077419  1174387  1280082  1395289  1520865  1657743  1806940  1969565  2146825  2340040  2550643  2780201  3030419  3303157  3600441  3924481  38250796
- OPERATING COSTS                     0        0        0        0        0        0        0   305309   521871   662728   773543   843162   919046  1001760  1091919  1190191  1297309  1414066  1541332  1680052  1831257  1996070  2175716  2371531  2584969  2817616  3071201  30090647
- FEDERAL INCOME TAX                  0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    61727   182821   199323   217307   236908   258270   281554   306930   1744840
- PROVINCIAL INCOME TAX               0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    15432    45705    49831    54327    59227    64568    70388    76733    436210
- INTEREST PAID FROM REVENUE          0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0         0
- DEBT REPAYMENTS                     0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0         0
- CAPITAL COSTS FROM REVENUE          0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0         0
- EQUITY CAPITAL COSTS                0        0   130664   427272   776211  1015284   922216   165391        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0   3437037
-------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
= CASH FLOW                           0        0  -130664  -427272  -776211 -1015284  -922216  -165391    60503   153428   214915   234258   255341   278322   303371   330674   360435   392874   428232   389615   280256   305420   332851   362754   395350   430883   469617   2542061
  ACCUMULATED                         0        0  -130664  -557936 -1334147 -2349430 -3271646 -3437037 -3376534 -3223106 -3008190 -2773933 -2518592 -2240270 -1936899 -1606225 -1245791  -852917  -424685   -35070   245186   550606   883457  1246210  1641561  2072444  2542061

DCF RATE OF RETURN (%)             4.33
PAYOUT PERIOD (YEARS)             13.13
EQUITY RATIO (%)                 100.00


*******************
COMPONENT INFLATION
*******************
PRICE          RATE                5.00    11.00    11.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00
               FACTOR             1.050    1.166    1.294    1.410    1.537    1.675    1.826    1.991    2.170    2.365    2.578    2.810    3.063    3.338    3.639    3.966    4.323    4.712    5.136    5.599    6.103    6.652    7.250    7.903    8.614    9.390   10.235

CAPITAL        RATE                5.00    11.00    11.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00
               FACTOR             1.050    1.166    1.294    1.410    1.537    1.675    1.826    1.991    2.170    2.365    2.578    2.810    3.063    3.338    3.639    3.966    4.323    4.712    5.136    5.599    6.103    6.652    7.250    7.903    8.614    9.390   10.235

OPERATING      RATE                5.00    11.00    11.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00
               FACTOR             1.050    1.166    1.294    1.410    1.537    1.675    1.826    1.991    2.170    2.365    2.578    2.810    3.063    3.338    3.639    3.966    4.323    4.712    5.136    5.599    6.103    6.652    7.250    7.903    8.614    9.390   10.235


********************
CAPITAL EXPENDITURES
********************
PROCESS FACILITIES                    0        0    86000   180000   300000   350000   264000        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0   1180000
BUILDINGS + SERVICES                  0        0        0        0        0    10000    10000        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     20000
INDIRECTS                             0        0    10000    63000   105000   126000   116000        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    420000
CONTINGENCY                           0        0     5000    60000   100000   120000   115000        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    400000
WORKING CAPITAL                       0        0        0        0        0        0        0    55490        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     55490
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TOTAL CAPITAL COSTS                   0        0   101000   303000   505000   606000   505000    55490        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0   2075490
INFLATION                             0        0    29664   124272   271211   409284   417216    54964        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0   1306611
CAPITALIZED COSTS                     0        0        0        0        0        0        0    54937        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     54937
-------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
TOTAL CAPITAL                         0        0   130664   427272   776211  1015284   922216   165391        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0   3437037
ACCUMULATED.                          0        0   130664   557936  1334147  2349430  3271646  3437037  3437037  3437037  3437037  3437037  3437037  3437037  3437037  3437037  3437037  3437037  3437037  3437037  3437037  3437037  3437037  3437037  3437037  3437037  3437037


***************
OPERATING COSTS
***************
OPERATING COSTS                       0        0        0        0        0        0        0   180980   240530   280230   300080   300080   300080   300080   300080   300080   300080   300080   300080   300080   300080   300080   300080   300080   300080   300080   300080   5803100
INFLATION                             0        0        0        0        0        0        0   179265   281341   382498   473463   543082   618966   701680   791839   890111   997229  1113986  1241252  1379972  1531177  1695990  1875636  2071451  2284889  2517536  2771121  24342484
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TOTAL OPERATING COSTS                 0        0        0        0        0        0        0   360245   521871   662728   773543   843162   919046  1001760  1091919  1190191  1297309  1414066  1541332  1680052  1831257  1996070  2175716  2371531  2584969  2817616  3071201  30145584
CAPITALIZED COSTS                     0        0        0        0        0        0        0    54937        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     54937
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TOTAL OPERATING COSTS                 0        0        0        0        0        0        0   305309   521871   662728   773543   843162   919046  1001760  1091919  1190191  1297309  1414066  1541332  1680052  1831257  1996070  2175716  2371531  2584969  2817616  3071201  30090647


******************
FEDERAL INCOME TAX
******************
+ REVENUE                             0        0        0        0        0        0        0   305309   582375   816156   988458  1077419  1174387  1280082  1395289  1520865  1657743  1806940  1969565  2146825  2340040  2550643  2780201  3030419  3303157  3600441  3924481  38250796
- OPERATING COSTS                     0        0        0        0        0        0        0   305309   521871   662728   773543   843162   919046  1001760  1091919  1190191  1297309  1414066  1541332  1680052  1831257  1996070  2175716  2371531  2584969  2817616  3071201  30090647
- INTEREST                            0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0         0
- CLASS 3                             0        0        0        0        0        0        0        0     1751     1663     1580     1501     1426     1355     1287     1223     1162     1103     1048      996      946      899      854      811      771      732      695     21882
- CLASS 29 CURRENT                    0        0        0        0        0        0        0        0    58753   151765        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    210510
- CLASS 29 CARRIED                    0        0        0        0        0        0        0        0        0        0   213335   232757   253915   276967   302084   329451   359273   391770   427184   294314        0        0        0        0        0        0        0   3081050
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= TAXABLE INCOME                      0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0   171464   507837   553675   603631   658077   717418   782093   852584   4846779
  INCOME TAX (36.0%)                  0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    61727   182821   199323   217307   236908   258270   281554   306930   1744840
- INVESTMENT TAX CREDIT               0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0         0
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= TAX PAYABLE AFTER ITC               0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    61727   182821   199323   217307   236908   258270   281554   306930   1744840


*********************
PROVINCIAL INCOME TAX
*********************
FEDERAL TAXABLE INCOME                0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0   171464   507837   553675   603631   658077   717418   782093   852584   4846779
INCOME TAX (9.0 %)                    0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    15432    45705    49831    54327    59227    64568    70388    76733    436210


CAPITAL COST ALLOWANCE
**********************
INPUT
  CLASS 3  - CCA                      0        0        0        0        0    16754    18262        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     35016
             ITC                      0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0         0
  CLASS 29 - CCA                      0        0   130664   427272   776211   998530   903954    54937        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0   3291567
             ITC                      0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0         0
  NO ALLOWANCE                        0        0        0        0        0        0        0   110454        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    110454
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  TOTAL INPUT                         0        0   130664   427272   776211  1015284   922216   165391        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0   3437037

POOLS BEFORE DEDUCTIONS
  CLASS 3                             0        0        0        0        0    16754    35016    35016    35016    33265    31602    30021    28520    27094    25740    24453    23230    22068    20965    19917    18921    17975    17076    16222    15411    14641    13909
  CLASS 29 CURRENT                    0        0        0        0        0        0        0   822092  1645784   822092        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0
  CLASS 29 CARRIED                    0        0        0        0        0        0        0   822092  2409923  3081050  3081050  2867715  2634958  2381043  2104076  1801992  1472541  1113268   721498   294314        0        0        0        0        0        0        0
```

Base Case
($50.00/tonne Coal)
April 13, 1982
Announcement
Figure 3.13

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                                                                                                                                  ($1000)

                                       1982     1983     1984     1985     1986     1987     1988     1989     1990     1991     1992     1993     1994     1995     1996     1997     1998     1999     2000     2001     2002     2003     2004     2005     2006     2007     2008       TOTAL
                                       ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----     ----       -----
**********************
CASH FLOW CALCULATIONS
**********************
+ REVENUE                             0        0        0        0        0        0        0   342124   652601   914573  1107650  1207338  1315999  1434439  1563538  1704257  1857640  2024827  2207062  2405697  2622210  2858209  3115448  3395838  3701463  4034595  4397709    42863215
- OPERATING COSTS                     0        0        0        0        0        0        0   237848   453696   635822   770050   839354   914896   997237  1086988  1184817  1291451  1407681  1534372  1672466  1822988  1987057  2165892  2360822  2573296  2804893  3057333    29798960
- FEDERAL INCOME TAX                  0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0   132455   221795   241810   263623   287396   313307   341548   372328   405877   442442   482297     3504879
- PROVINCIAL INCOME TAX               0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    33114    55449    60452    65906    71849    78327    85387    93082   101469   110611   120574      876220
- INTEREST PAID FROM REVENUE          0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0           0
- DEBT REPAYMENTS                     0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0           0
- CAPITAL COSTS FROM REVENUE          0        0        0        0        0        0        0   104276        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      104276
- EQUITY CAPITAL COSTS                0        0   130664   437143   791581  1032037   949608    17425        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     3358459
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= CASH FLOW                           0        0  -130664  -437143  -791581 -1032037  -949608   -17425   198905   278751   337600   367984   401103   437202   476550   519440   408621   339902   370427   403703   439977   479518   522621   569605   620821   676649   737504     5220422
  ACCUMULATED                         0        0  -130664  -567807 -1359388 -2391426 -3341034 -3358459 -3159555 -2880803 -2543203 -2175219 -1774117 -1336915  -860365  -348925    59695   399597   770025  1173728  1613704  2093222  2615843  3185448  3806269  4482919  5220423

DCF RATE OF RETURN (%)             7.74
PAYOUT PERIOD (YEARS)              9.85
EQUITY RATIO (%)                 100.00


********************
COMPONENT INFLATION
********************
PRICE         RATE                 5.00    11.00    11.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00
              FACTOR              1.050    1.166    1.294    1.410    1.537    1.675    1.826    1.991    2.170    2.365    2.578    2.810    3.063    3.338    3.639    3.966    4.323    4.712    5.136    5.599    6.103    6.652    7.250    7.903    8.614    9.390   10.235

CAPITAL       RATE                 5.00    11.00    11.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00
              FACTOR              1.050    1.166    1.294    1.410    1.537    1.675    1.826    1.991    2.170    2.365    2.578    2.810    3.063    3.338    3.639    3.966    4.323    4.712    5.136    5.599    6.103    6.652    7.250    7.903    8.614    9.390   10.235

OPERATING     RATE                 5.00    11.00    11.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00     9.00
              FACTOR              1.050    1.166    1.294    1.410    1.537    1.675    1.826    1.991    2.170    2.365    2.578    2.810    3.063    3.338    3.639    3.966    4.323    4.712    5.136    5.599    6.103    6.652    7.250    7.903    8.614    9.390   10.235


********************
CAPITAL EXPENDITURES
********************
PROCESS FACILITIES                    0        0    86000   185000   305000   355000   271160        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     1202160
BUILDINGS + SERVICES                  0        0        0        0        0    10000    10000        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0       20000
INDIRECTS                             0        0    10000    63000   107000   128000   119740        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      427740
CONTINGENCY                           0        0     5000    62000   103000   123000   119100        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      412100
WORKING CAPITAL                       0        0        0        0        0        0        0    61140        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0       61140
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TOTAL CAPITAL COSTS                   0        0   101000   310000   515000   616000   520000    61140        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     2123140
INFLATION                             0        0    29664   127143   276581   416037   429608    60561        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     1339595
CAPITALIZED COSTS                     0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0           0
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TOTAL CAPITAL                         0        0   130664   437143   791581  1032037   949608   121701        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     3462735
ACCUMULATED                           0        0   130664   567807  1359388  2391426  3341034  3462735  3462735  3462735  3462735  3462735  3462735  3462735  3462735  3462735  3462735  3462735  3462735  3462735  3462735  3462735  3462735  3462735  3462735  3462735  3462735


***************
OPERATING COSTS
***************
OPERATING COSTS                       0        0        0        0        0        0        0   119490   209108   268853   298725   298725   298725   298725   298725   298725   298725   298725   298725   298725   298725   298725   298725   298725   298725   298725   298725     5675776
INFLATION                             0        0        0        0        0        0        0   118358   244588   366969   471325   540629   616171   698512   788263   886092   992726  1108956  1235647  1373741  1524263  1688332  1867167  2062097  2274571  2506168  2758608    24123184
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TOTAL OPERATING COSTS                 0        0        0        0        0        0        0   237848   453696   635822   770050   839354   914896   997237  1086988  1184817  1291451  1407681  1534372  1672466  1822988  1987057  2165892  2360822  2573296  2804893  3057333    29798960
CAPITALIZED COSTS                     0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0           0
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TOTAL OPERATING COSTS                 0        0        0        0        0        0        0   237848   453696   635822   770050   839354   914896   997237  1086988  1184817  1291451  1407681  1534372  1672466  1822988  1987057  2165892  2360822  2573296  2804893  3057333    29798960


******************
FEDERAL INCOME TAX
******************
+ REVENUE                             0        0        0        0        0        0        0   342124   652601   914573  1107650  1207338  1315999  1434439  1563538  1704257  1857640  2024827  2207062  2405697  2622210  2858209  3115448  3395838  3701463  4034595  4397709    42863215
- OPERATING COSTS                     0        0        0        0        0        0        0   237848   453696   635822   770050   839354   914896   997237  1086988  1184817  1291451  1407681  1534372  1672466  1822988  1987057  2165892  2360822  2573296  2804893  3057333    29798960
- INTEREST                            0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0           0
- CLASS 3                             0        0        0        0        0        0        0     1751     1663     1580     1501     1426     1355     1287     1223     1162     1103     1048      996      946      899      854      811      771      732      695      661       22463
- CLASS 29 CURRENT                    0        0        0        0        0        0        0   102525   197242   277171        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      576937
- CLASS 29 CARRIED                    0        0        0        0        0        0        0        0        0        0   336099   366558   399748   435915   475327   518278   197156        0        0        0        0        0        0        0        0        0        0     2729081
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= TAXABLE INCOME                      0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0   367929   616098   671693   732285   798323   870298   948745  1034245  1127435  1229007  1339715     9735774
  INCOME TAX (36.0%)                  0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0   132455   221795   241810   263623   287396   313307   341548   372328   405877   442442   482297     3504879
- INVESTMENT TAX CREDIT               0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0           0
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= TAX PAYABLE AFTER ITC               0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0   132455   221795   241810   263623   287396   313307   341548   372328   405877   442442   482297     3504879


*********************
PROVINCIAL INCOME TAX
*********************
FEDERAL TAXABLE INCOME                0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0   367929   616098   671693   732285   798323   870298   948745  1034245  1127435  1229007  1339715     9735774
INCOME TAX (9.0 %)                    0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0    33114    55449    60452    65906    71849    78327    85387    93082   101469   110611   120574      876220

CAPITAL COST ALLOWANCE
***********************
INPUT
   CLASS 3  - CCA                     0        0        0        0        0    16754    18262        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0       35016
              ITC                     0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0           0
   CLASS 29 - CCA                     0        0   130664   437143   791581  1015284   931347        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     3306019
              ITC                     0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0           0
   NO ALLOWANCE                       0        0        0        0        0        0        0   121701        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0      121701
----------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------
   TOTAL INPUT                        0        0   130664   437143   791581  1032037   949608   121701        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0     3462735

POOLS BEFORE DEDUCTIONS
   CLASS 3                            0        0        0        0        0    16754    35016    35016    33265    31602    30021    28520    27094    25740    24453    23230    22068    20965    19917    18921    17975    17076    16222    15411    14641    13909    13213
   CLASS 29 CURRENT                   0        0        0        0        0        0        0   826505  1653009   826505        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0        0
   CLASS 29 CARRIED                   0        0        0        0        0        0        0   723980  2179748  2729081  2729081  2392982  2026424  1626676  1190762   715434   197156        0        0        0        0        0        0        0        0        0        0
```

Alternate Case 1
($50.00/tonne Coal)
April 13, 1982 Agreements
Figure 3.14

($1000)

CASH FLOW CALCULATIONS

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
+ REVENUE	0	0	0	0	0	0	0	388302	740685	1038019	1257154	1370298	1493624	1628051	1774575	1934287	2108373	2298126	2504958	2730404	2976140	3243993	3535952	3854188	4201065	4579161	4991285	48648639
- OPERATING COSTS	0	0	0	0	0	0	0	286373	546256	765539	927153	1010597	1101551	1200690	1308752	1426540	1554929	1694872	1847411	2013678	2194909	2392450	2607771	2842470	3098293	3377139	3681081	35878453
- FEDERAL INCOME TAX	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	37498	236287	257613	280855	306186	333795	363885	396682	432427	471388	3116617
- PROVINCIAL INCOME TAX	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	9374	59072	64403	70214	76547	83449	90971	99170	108107	117847	779154
- INTEREST PAID FROM REVENUE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
- DEBT REPAYMENTS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
- CAPITAL COSTS FROM REVENUE	0	0	0	0	0	0	0	101929	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	101929
- EQUITY CAPITAL COSTS	0	0	139720	479447	884880	1206278	1417656	36781	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	4164762
= CASH FLOW	0	0	-139720	-479447	-884880	-1206278	-1417656	-36781	194429	272479	330001	359701	392074	427361	465823	507747	553444	556382	362189	394710	430163	468809	510938	556861	606920	661488	720969	4607725
ACCUMULATED	0	0	-139720	-619167	-1504047	-2710325	-4127981	-4164762	-3970333	-3697854	-3367853	-3008152	-2616078	-2188718	-1722895	-1215148	-661704	-105322	256867	651577	1081740	1550550	2061487	2618348	3225268	3886756	4607725	

DCF RATE OF RETURN (%) 6.07
PAYOUT PERIOD (YEARS) 11.29
EQUITY RATIO (%) 100.00

COMPONENT INFLATION

		1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008
PRICE	RATE	5.00	11.00	11.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00
	FACTOR	1.050	1.166	1.294	1.410	1.537	1.675	1.826	1.991	2.170	2.365	2.578	2.810	3.063	3.338	3.639	3.966	4.323	4.712	5.136	5.599	6.103	6.652	7.250	7.903	8.614	9.390	10.235
CAPITAL	RATE	5.00	11.00	11.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00
	FACTOR	1.050	1.166	1.294	1.410	1.537	1.675	1.826	1.991	2.170	2.365	2.578	2.810	3.063	3.338	3.639	3.966	4.323	4.712	5.136	5.599	6.103	6.652	7.250	7.903	8.614	9.390	10.235
OPERATING	RATE	5.00	11.00	11.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00
	FACTOR	1.050	1.166	1.294	1.410	1.537	1.675	1.826	1.991	2.170	2.365	2.578	2.810	3.063	3.338	3.639	3.966	4.323	4.712	5.136	5.599	6.103	6.652	7.250	7.903	8.614	9.390	10.235

CAPITAL EXPENDITURES

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
PROCESS FACILITIES	0	0	90000	200000	320000	400000	459700	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1469700
BUILDINGS + SERVICES	0	0	0	0	0	12000	12000	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	24000
INDIRECTS	0	0	12000	70000	130700	157000	153100	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	522800
CONTINGENCY	0	0	6000	70000	125000	151000	151500	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	503500
WORKING CAPITAL	0	0	0	0	0	0	0	69685	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	69685
TOTAL CAPITAL COSTS	0	0	108000	340000	575700	720000	776300	69685	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	2589685
INFLATION	0	0	31720	139447	309180	486278	641356	69025	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1677006
CAPITALIZED COSTS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
TOTAL CAPITAL	0	0	139720	479447	884880	1206278	1417656	138710	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	4266691
ACCUMULATED	0	0	139720	619167	1504047	2710325	4127981	4266691	4266691	4266691	4266691	4266691	4266691	4266691	4266691	4266691	4266691	4266691	4266691	4266691	4266691	4266691	4266691	4266691	4266691	4266691	4266691	

OPERATING COSTS

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
OPERATING COSTS	0	0	0	0	0	0	0	143868	251769	323703	359670	359670	359670	359670	359670	359670	359670	359670	359670	359670	359670	359670	359670	359670	359670	359670	359670	6833730
INFLATION	0	0	0	0	0	0	0	142505	294487	441836	567483	650927	741881	841020	949082	1066870	1195259	1335202	1487741	1654008	1835239	2032780	2248101	2482800	2738623	3017469	3321411	29044723
TOTAL OPERATING COSTS	0	0	0	0	0	0	0	286373	546256	765539	927153	1010597	1101551	1200690	1308752	1426540	1554929	1694872	1847411	2013678	2194909	2392450	2607771	2842470	3098293	3377139	3681081	35878453
CAPITALIZED COSTS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
TOTAL OPERATING COSTS	0	0	0	0	0	0	0	286373	546256	765539	927153	1010597	1101551	1200690	1308752	1426540	1554929	1694872	1847411	2013678	2194909	2392450	2607771	2842470	3098293	3377139	3681081	35878453

FEDERAL INCOME TAX

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
+ REVENUE	0	0	0	0	0	0	0	388302	740685	1038019	1257154	1370298	1493624	1628051	1774575	1934287	2108373	2298126	2504958	2730404	2976140	3243993	3535952	3854188	4201065	4579161	4991285	48648639
- OPERATING COSTS	0	0	0	0	0	0	0	286373	546256	765539	927153	1010597	1101551	1200690	1308752	1426540	1554929	1694872	1847411	2013678	2194909	2392450	2607771	2842470	3098293	3377139	3681081	35878453
- INTEREST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
- CLASS 3	0	0	0	0	0	0	0	2101	1996	1896	1801	1711	1626	1544	1467	1394	1324	1258	1195	1135	1078	1025	973	925	878	835	793	26956
- CLASS 29 CURRENT	0	0	0	0	0	0	0	99828	192433	270583	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	562844
- CLASS 29 CARRIED	0	0	0	0	0	0	0	0	0	0	328199	357990	390448	425816	464356	506353	552120	497836	0	0	0	0	0	0	0	0	0	3523118
= TAXABLE INCOME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	104160	656352	715591	780153	850518	927208	1010793	1101894	1201187	1309411	8657268
INCOME TAX (36.0%)	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	37498	236287	257613	280855	306186	333795	363885	396682	432427	471388	3116617
- INVESTMENT TAX CREDIT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
= TAX PAYABLE AFTER ITC	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	37498	236287	257613	280855	306186	333795	363885	396682	432427	471388	3116617

PROVINCIAL INCOME TAX

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
FEDERAL TAXABLE INCOME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	104160	656352	715591	780153	850518	927208	1010793	1101894	1201187	1309411	8657268
INCOME TAX (9.0 %)	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	9374	59072	64403	70214	76547	83449	90971	99170	108107	117847	779154

CAPITAL COST ALLOWANCE

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
INPUT																												
CLASS 3 - CCA	0	0	0	0	0	20105	21914	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	42019
ITC	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CLASS 29 - CCA	0	0	139720	479447	884880	1186173	1395742	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	4085962
ITC	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO ALLOWANCE	0	0	0	0	0	0	0	138710	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	138710
TOTAL INPUT	0	0	139720	479447	884880	1206278	1417656	138710	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	4266691
POOLS BEFORE DEDUCTIONS																												
CLASS 3	0	0	0	0	0	20105	42019	42019	39918	37922	36025	34224	32513	30888	29345	27876	26482	25158	23900	22705	21570	20491	19467	18494	17569	16690	15856	
CLASS 29 CURRENT	0	0	0	0	0	0	0	1021491	2042981	1021491	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	
CLASS 29 CARRIED	0	0	0	0	0	0	0	921663	2772211	3523118	3523118	3194919	2836929	2446481	2020665	1556309	1049956	497836	0	0	0	0	0	0	0	0	0	

Alternate Case 2
($50.00/tonne Coal)
April 13, 1982
Announcement
Figure 3.15

($1000)

CASH FLOW CALCULATIONS

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
+ REVENUE	0	0	0	0	0	0	0	412544	786932	1102827	1335647	1455856	1586883	1729702	1885375	2055059	2240015	2441616	2661361	2900884	3161963	3446540	3756729	4094834	4463369	4865072	5302929	51686139
- OPERATING COSTS	0	0	0	0	0	0	0	302114	576282	807619	978116	1066146	1162100	1266689	1380691	1504953	1640398	1788034	1948957	2124364	2315556	2523956	2751112	2998712	3268597	3562770	3883420	37850586
- FEDERAL INCOME TAX	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	94282	256035	279139	304318	331761	361671	394271	429802	468528	510738	3430546
- PROVINCIAL INCOME TAX	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	23571	64009	69785	76080	82940	90418	98568	107450	117132	127684	857637
- INTEREST PAID FROM REVENUE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
- DEBT REPAYMENTS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
- CAPITAL COSTS FROM REVENUE	0	0	0	0	0	0	0	110430	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	110430
- EQUITY CAPITAL COSTS	0	0	139720	479447	884880	1290047	1527226	35595	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	4356915
= CASH FLOW	0	0	-139720	-479447	-884880	-1290047	-1527226	-35595	210649	295209	357531	389709	424783	463014	504685	550107	599616	535729	392360	427597	466009	507882	553527	603283	657520	716642	781087	5080024
ACCUMULATED	0	0	-139720	-619167	-1504047	-2794094	-4321320	-4356915	-4146266	-3851057	-3493525	-3103816	-2679033	-2216019	-1711334	-1161228	-561612	-25883	366477	794074	1260083	1767965	2321492	2924775	3582295	4298937	5080024	

DCF RATE OF RETURN (%)	6.35
PAYOUT PERIOD (YEARS)	11.07
EQUITY RATIO (%)	100.00

COMPONENT INFLATION

		1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008
PRICE	RATE	5.00	11.00	11.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00
	FACTOR	1.050	1.166	1.294	1.410	1.537	1.675	1.826	1.991	2.170	2.365	2.578	2.810	3.063	3.338	3.639	3.966	4.323	4.712	5.136	5.599	6.103	6.652	7.250	7.903	8.614	9.390	10.235
CAPITAL	RATE	5.00	11.00	11.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00
	FACTOR	1.050	1.166	1.294	1.410	1.537	1.675	1.826	1.991	2.170	2.365	2.578	2.810	3.063	3.338	3.639	3.966	4.323	4.712	5.136	5.599	6.103	6.652	7.250	7.903	8.614	9.390	10.235
OPERATING	RATE	5.00	11.00	11.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00
	FACTOR	1.050	1.166	1.294	1.410	1.537	1.675	1.826	1.991	2.170	2.365	2.578	2.810	3.063	3.338	3.639	3.966	4.323	4.712	5.136	5.599	6.103	6.652	7.250	7.903	8.614	9.390	10.235

CAPITAL EXPENDITURES

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
PROCESS FACILITIES	0	0	90000	200000	320000	420000	505405	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1535405
BUILDINGS + SERVICES	0	0	0	0	0	12000	12000	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	24000
INDIRECTS	0	0	12000	70000	130700	170000	162400	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	545100
CONTINGENCY	0	0	6000	70000	125000	168000	156495	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	525495
WORKING CAPITAL	0	0	0	0	0	0	0	73360	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	73360
TOTAL CAPITAL COSTS	0	0	108000	340000	575700	770000	836300	73360	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	2703360
INFLATION	0	0	31720	139447	309180	520047	690926	72665	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1763985
CAPITALIZED COSTS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
TOTAL CAPITAL	0	0	139720	479447	884880	1290047	1527226	146025	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	4467345
ACCUMULATED	0	0	139720	619167	1504047	2794094	4321320	4467345	4467345	4467345	4467345	4467345	4467345	4467345	4467345	4467345	4467345	4467345	4467345	4467345	4467345	4467345	4467345	4467345	4467345	4467345	4467345	

OPERATING COSTS

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
OPERATING COSTS	0	0	0	0	0	0	0	151776	265608	341496	379440	379440	379440	379440	379440	379440	379440	379440	379440	379440	379440	379440	379440	379440	379440	379440	379440	7209360
INFLATION	0	0	0	0	0	0	0	150338	310674	466123	598676	686706	782660	887249	1001251	1125513	1260958	1408594	1569517	1744924	1936116	2144516	2371672	2619272	2889157	3183330	3503980	30641226
TOTAL OPERATING COSTS	0	0	0	0	0	0	0	302114	576282	807619	978116	1066146	1162100	1266689	1380691	1504953	1640398	1788034	1948957	2124364	2315556	2523956	2751112	2998712	3268597	3562770	3883420	37850586
CAPITALIZED COSTS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
TOTAL OPERATING COSTS	0	0	0	0	0	0	0	302114	576282	807619	978116	1066146	1162100	1266689	1380691	1504953	1640398	1788034	1948957	2124364	2315556	2523956	2751112	2998712	3268597	3562770	3883420	37850586

FEDERAL INCOME TAX

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
+ REVENUE	0	0	0	0	0	0	0	412544	786932	1102827	1335647	1455856	1586883	1729702	1885375	2055059	2240015	2441616	2661361	2900884	3161963	3446540	3756729	4094834	4463369	4865072	5302929	51686139
- OPERATING COSTS	0	0	0	0	0	0	0	302114	576282	807619	978116	1066146	1162100	1266689	1380691	1504953	1640398	1788034	1948957	2124364	2315556	2523956	2751112	2998712	3268597	3562770	3883420	37850586
- INTEREST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
- CLASS 3	0	0	0	0	0	0	0	2101	1996	1896	1801	1711	1626	1544	1467	1394	1324	1258	1195	1135	1078	1025	973	925	878	835	793	26956
- CLASS 29 CURRENT	0	0	0	0	0	0	0	108329	208653	293313	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	610295
- CLASS 29 CARRIED	0	0	0	0	0	0	0	0	0	0	355730	387998	423158	461469	503218	548713	598292	390429	0	0	0	0	0	0	0	0	0	3669006
= TAXABLE INCOME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	261895	711209	775385	845329	921559	1004643	1095197	1193894	1301468	1418717	9529295
INCOME TAX (36.0%)	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	94282	256035	279139	304318	331761	361671	394271	429802	468528	510738	3430546
- INVESTMENT TAX CREDIT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
= TAX PAYABLE AFTER ITC	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	94282	256035	279139	304318	331761	361671	394271	429802	468528	510738	3430546

PROVINCIAL INCOME TAX

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
FEDERAL TAXABLE INCOME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	261895	711209	775385	845329	921559	1004643	1095197	1193894	1301468	1418717	9529295
INCOME TAX (9.0 %)	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	23571	64009	69785	76080	82940	90418	98568	107450	117132	127684	857637

CAPITAL COST ALLOWANCE

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
INPUT																												
CLASS 3 - CCA	0	0	0	0	0	20105	21914	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	42019
ITC	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CLASS 29 - CCA	0	0	139720	479447	884880	1269942	1505312	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	4279302
ITC	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO ALLOWANCE	0	0	0	0	0	0	0	146025	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	146025
TOTAL INPUT	0	0	139720	479447	884880	1290047	1527226	146025	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	4467345
POOLS BEFORE DEDUCTIONS																												
CLASS 3	0	0	0	0	0	20105	42019	42019	39918	37922	36026	34224	32513	30888	29343	27876	26482	25158	23900	22705	21570	20491	19467	18494	17569	16691	15856	
CLASS 29 CURRENT	0	0	0	0	0	0	0	1069825	2139651	1069825	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	
CLASS 29 CARRIED	0	0	0	0	0	0	0	961496	2092494	3669006	3669006	3313276	2925278	2502120	2040651	1537433	988721	390429	0	0	0	0	0	0	0	0	0	

Alternate Case 3
($50.00/tonne Coal)
April 13, 1982
Announcement
Figure 3.16

($1000)

CASH FLOW CALCULATIONS

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
+ REVENUE	0	0	0	0	0	0	0	493017	940432	1317947	1596179	1739836	1896421	2067099	2253137	2455920	2676953	2917878	3180487	3466731	3778737	4118823	4489518	4893574	5333996	5814055	6337320	61760061
- OPERATING COSTS	0	0	0	0	0	0	0	391449	746689	1046431	1267344	1381405	1505731	1641247	1788959	1949966	2125462	2316754	2525262	2752536	3000264	3270287	3564613	3885429	4235117	4616278	5031743	49042964
- FEDERAL INCOME TAX	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	237230	279862	305104	332615	362600	395280	430899	469723	2813314
- PROVINCIAL INCOME TAX	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	59308	69966	76276	83154	90650	98820	107725	117431	703328
- INTEREST PAID FROM REVENUE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
- DEBT REPAYMENTS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
- CAPITAL COSTS FROM REVENUE	0	0	0	0	0	0	0	101569	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	101569
- EQUITY CAPITAL COSTS	0	0	194961	637524	1158168	1534988	1399759	59465	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	4984865
= CASH FLOW	0	0	-194961	-637524	-1158168	-1534988	-1399759	-59465	193743	271517	328836	358431	390690	425852	464178	505954	551490	601124	655225	417658	428646	467156	509135	554896	604777	659153	718425	4122021
ACCUMULATED	0	0	-194961	-832485	-1990653	-3525641	-4925400	-4984865	-4791122	-4519606	-4190770	-3832339	-3441650	-3015798	-2551620	-2045666	-1494175	-893051	-237826	179832	608478	1075633	1584769	2139665	2744444	3403597	4122021	

DCF RATE OF RETURN (%)	4.78
PAYOUT PERIOD (YEARS)	12.57
EQUITY RATIO (%)	100.00

COMPONENT INFLATION

		1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008
PRICE	RATE	5.00	11.00	11.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00
	FACTOR	1.050	1.166	1.294	1.410	1.537	1.675	1.826	1.991	2.170	2.365	2.578	2.810	3.063	3.338	3.639	3.966	4.323	4.712	5.136	5.599	6.103	6.652	7.250	7.903	8.614	9.390	10.235
CAPITAL	RATE	5.00	11.00	11.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00
	FACTOR	1.050	1.166	1.294	1.410	1.537	1.675	1.826	1.991	2.170	2.365	2.578	2.810	3.063	3.338	3.639	3.966	4.323	4.712	5.136	5.599	6.103	6.652	7.250	7.903	8.614	9.390	10.235
OPERATING	RATE	5.00	11.00	11.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00	9.00
	FACTOR	1.050	1.166	1.294	1.410	1.537	1.675	1.826	1.991	2.170	2.365	2.578	2.810	3.063	3.338	3.639	3.966	4.323	4.712	5.136	5.599	6.103	6.652	7.250	7.903	8.614	9.390	10.235

CAPITAL EXPENDITURES

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
PROCESS FACILITIES	0	0	88800	266400	444000	532800	444810	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1776810
BUILDINGS + SERVICES	0	0	0	0	0	12000	12000	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	24000
INDIRECTS	0	0	31500	94500	157500	189000	157690	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	630190
CONTINGENCY	0	0	30400	91200	152000	182400	152000	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	608000
WORKING CAPITAL	0	0	0	0	0	0	0	80900	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	80900
TOTAL CAPITAL COSTS	0	0	150700	452100	753500	916200	766500	80900	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	3119900
INFLATION	0	0	44261	185424	404668	618788	633259	80134	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1966534
CAPITALIZED COSTS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
TOTAL CAPITAL	0	0	194961	637524	1158168	1534988	1399759	161034	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	5086434
ACCUMULATED	0	0	194961	832485	1990653	3525641	4925400	5086434	5086434	5086434	5086434	5086434	5086434	5086434	5086434	5086434	5086434	5086434	5086434	5086434	5086434	5086434	5086434	5086434	5086434	5086434	5086434	

OPERATING COSTS

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
OPERATING COSTS	0	0	0	0	0	0	0	196656	344148	442476	491640	491640	491640	491640	491640	491640	491640	491640	491640	491640	491640	491640	491640	491640	491640	491640	491640	9341160
INFLATION	0	0	0	0	0	0	0	194793	402541	603955	775704	889765	1014091	1149607	1297319	1458326	1633822	1825114	2033622	2260896	2508624	2778647	3072973	3393789	3743477	4124638	4540103	39701804
TOTAL OPERATING COSTS	0	0	0	0	0	0	0	391449	746689	1046431	1267344	1381405	1505731	1641247	1788959	1949966	2125462	2316754	2525262	2752536	3000264	3270287	3564613	3885429	4235117	4616278	5031743	49042964
CAPITALIZED COSTS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
TOTAL OPERATING COSTS	0	0	0	0	0	0	0	391449	746689	1046431	1267344	1381405	1505731	1641247	1788959	1949966	2125462	2316754	2525262	2752536	3000264	3270287	3564613	3885429	4235117	4616278	5031743	49042964

FEDERAL INCOME TAX

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
+ REVENUE	0	0	0	0	0	0	0	493017	940432	1317947	1596179	1739836	1896421	2067099	2253137	2455920	2676953	2917878	3180487	3466731	3778737	4118823	4489518	4893574	5333996	5814055	6337320	61760061
- OPERATING COSTS	0	0	0	0	0	0	0	391449	746689	1046431	1267344	1381405	1505731	1641247	1788959	1949966	2125462	2316754	2525262	2752536	3000264	3270287	3564613	3885429	4235117	4616278	5031743	49042964
- INTEREST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
- CLASS 3	0	0	0	0	0	0	0	2101	1996	1896	1801	1711	1626	1544	1467	1394	1324	1258	1195	1135	1078	1025	973	925	879	835	793	26956
- CLASS 29 CURRENT	0	0	0	0	0	0	0	99468	191747	269621	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	560836
- CLASS 29 CARRIED	0	0	0	0	0	0	0	0	0	0	327034	356720	389064	424308	462711	504560	550166	599866	654030	54087	0	0	0	0	0	0	0	4322546
= TAXABLE INCOME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	658973	777395	847511	923931	1007221	1098000	1196943	1304785	7814768
INCOME TAX (36.0%)	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	237230	279862	305104	332615	362600	395280	430899	469723	2813314
- INVESTMENT TAX CREDIT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
= TAX PAYABLE AFTER ITC	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	237230	279862	305104	332615	362600	395280	430899	469723	2813314

PROVINCIAL INCOME TAX

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
FEDERAL TAXABLE INCOME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	658973	777395	847511	923931	1007221	1098000	1196943	1304785	7814768
INCOME TAX (9.0 %)	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	59308	69966	76276	83154	90650	98820	107725	117431	703328

CAPITAL COST ALLOWANCE

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	TOTAL
INPUT																												
CLASS 3 - CCA	0	0	0	0	0	20105	21914	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	42019
ITC	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CLASS 29 - CCA	0	0	194961	637524	1158168	1514883	1377845	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	4883381
ITC	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO ALLOWANCE	0	0	0	0	0	0	0	161034	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	161034
TOTAL INPUT	0	0	194961	637524	1158168	1534988	1399759	161034	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	5086434
POOLS BEFORE DEDUCTIONS																												
CLASS 3	0	0	0	0	0	20105	42019	42019	39918	37922	36026	34224	32513	30888	29343	27876	26482	25158	23900	22705	21570	20491	19467	18494	17569	16690	15856	
CLASS 29 CURRENT	0	0	0	0	0	0	0	1220845	2441691	1220846	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	
CLASS 29 CARRIED	0	0	0	0	0	0	0	1121378	3371321	4322546	4322546	3995512	3638792	3249729	2825421	2362710	1858150	1307984	708117	54087	0	0	0	0	0	0	0	

Alternate Case 4
($50.00/tonne Coal)
April 13, 1982
Figure 3.17

Volume

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REQ. NO. 1-9030

CONTRACT REPORT NO. OSQ81-00057

Western Canada Coal Liquefaction Study
Prefeasibility Study

Prepared for:

British Columbia Resources Investment Corporation
Petro-Canada Exploration Incorporated
Westcoast Transmission Company Limited
British Columbia Coal Limited
Energy, Mines and Resources Canada

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SBRG CONTINUOUS TESTS

HRI MICROAUTOCLAVE TESTS

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BCR Microautoclave Tests

PROJECT REPORT

BENCH SCALE EVALUATION OF TWO B.C. COALS BY TUBING BOMB LIQUEFACTION TESTS

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SUMMARY

Samples of coal from two B.C. deposits - "Greenhills" (B.C. Coal Ltd.) and "Monkman" (Petro-Canada) were tested in a tubing bomb apparatus to determine yields of selected products (gases, distillate, "soluble" and residue) under conditions modelling selected direct liquefaction processes.

A sample of Illinois #6 coal was also tested for comparison purposes.

For test conditions used reactivities of the two B.C. coals appear to be similar. However, under laboratory best yield conditions, both B.C. caols produced significantly lower amounts of distillate than Illinois #6 coal. Best laboratory yield conditions for the Illinois #6 coal and the B.C. coals were significantly different.

Test data indicate that the process types modelled can be ranked in order of increasing distillate yield as follows:

Donor Solvent	<	H-Coal	<	Bergius
(hydrogenated anthracene oil, no catalyst in liquefaction stage)		(hydrogenated anthracene oil, HDS1442A catalyst)		(Bergbau-Forschung solvent, red mud catalyst)

For Illinois #6 coal/H-Coal model tests, laboratory best yields of distillate were significantly higher than (published) yields from pilot plant tests on the same coal. Consequently, laboratory best yield results are considered to be valid for qualitative process comparison but not for quantitative projection of steady state yield.

Under laboratory best (distillate) yield conditions, both B.C. coals produced substantial amounts of carbon disulphide (CS_2) soluble, non-distillable ("pitch") material which was resistant to hydrogenation with either HDS - or red mud catalysts. Development of economic methods for utilizing this pitch byproduct (e.g. by upgrading, direct marketing) would be desirable to improve overall process yields.

INTRODUCTION

BACKGROUND

On behalf of a consortium, Kilborn Ltd. is performing a pre-feasibility study on direct liquefaction of selected B.C. coals by methods developed to at least small pilot scale.

Following an exchange of correspondence, Kilborn Ltd. awarded B.C. Research a subcontract to undertake bench scale "tubing bomb" liquefaction tests to provide initial data on conversion and product quality under specific conditions. General terms and conditions of the subcontract were outlined in a letter and attachment from Dr. N.E. Anderson of Kilborn Ltd. dated July 7, 1981.

Following project authorization, a planning meeting was held at B.C. Research. Representatives of Kilborn Ltd., Sandwell Beak Research Group, a consortium member and B.C. Research agreed to an initial experimental plan. Communication of initial results, by telephone conference of the above representatives (see minutes of meeting, Appendix 1) resulted in revisions to the initial plan as outlined below.

OBJECTIVES

The revised objectives of the study were:

- to model selected process conditions as closely as possible using the tubing bomb apparatus and three selected coals: Greenhills, Monkman and Illinois No. 6 (reference for reactivity).

- to determine conversion yields of specified products (gases, distillate fractions, residue).

- to report experimental data with evaluation of experimental accuracy.

MATERIALS AND METHODS

COAL SAMPLES

Four drums of Greenhills coal were received from B.C. Coal Ltd. Two drums contained sample of No. 1 seam, the third drum a sample of No. 7 seam and the fourth a sample of No. 10 seam. Core samples were taken from each drum and crushed to 10 mesh. A blend of the samples of seams No 1, 7 and 10 was prepared in the 2:1:1 proportion. A 1 kg split of this blend was ground to pass 70 mesh.

Monkman coal (Petro-Canada) was received as a sample of approximately 15 kg. It was crushed to -10 mesh, split and a 1 kg portion was ground to pass 70 mesh.

Illinois No. 6 coal was received as -70 mesh sample.

SOLVENTS

Anthracene oil (AO) was received from Sandwell Beak Laboratories and was used directly.

Hydrogenated Anthracene Oil (HAO) was prepared by catalytic hydrogenation of AO (see below).

Bergbau-Forschung Recycle oil (BF) was used directly.

CATALYSTS

Cyanamid HDS 1442-A catalyst was obtained from Hydrocarbon Research Institute and was ground to pass 70 mesh. (Analysis: Appendix 3)

A red mud (RM) sample provided by Kilborn Ltd. was prepared by filtration, drying, crushing and screening to -70 mesh.

HYDROGENATION OF ANTHRACENE OIL

The anthracene oil obtained from Sandwell-Beak was hydrogenated in a 2 ℓ autoclave on HDS-1442A catalyst. The conditions were as follows:

Charge:	1250 g of anthracene oil 50 g of HDS catalyst pellets 3 g of sulfur
Initial cold presure:	1700 psig of hydrogen
Heat up time:	2 hours
Temperature:	325°C
Maximum pressure at temperature:	2680 psig
Time at temperature:	1 hour

The product was pumped out of the reactor and filtered through No. 1 Whatman filter paper.

TUBING BOMB TESTS

The tubing bomb apparatus (Figure 1) was used as follows:

Charge:	coal (air dry) (specified wt) solvent (specified wt) catalyst and sulphur (specified wt) 2 x 3/8 in. steel balls for agitation
Closure:	screw type (soft copper gasket)
Flushing:	1 x N_2 2 x H_2
Pressurization:	cylinder H_2 to 1000 psig at ambient temperature unless otherwise specified
Agitation:	shaker

Heating:	propane fired fluid bed sand bath; 3-4 minutes to operating temperature
Temperature control:	manual in response to internal thermocouple (normally ± 3°C)
Cooling:	water spray (~3 min to T ≤300°C
Reactor Pressure:	gauge readings (±25 psig)
Gas Composition:	pressurized sample of reaction gas phase analyzed by gas chromatography

SLURRY WORK-UP PROCEDURE AND CALCULATIONS

The contents of the bombs were washed out with CS_2, filtered through tared soxhlet thimbles and extracted with CS_2 in the soxhlet apparatus until clear. A sample of CS_2 solution was taken for determination of +525°C residue by thermogravimetric analysis and for simulated distillation by gas chromatography when required (see below).

Thimbles containing undissolved residue were dried in a vacuum oven at 110°C to constant weight.

Total conversion was calculated from the residue weight and the known proximate analysis of the coal:

$$\text{Conversion (wt. \% DAF coal)} = \frac{(\text{Dry coal charge})-(CS_2\,\text{insolubles}) \times 100}{\text{DAF coal}}$$

In the above expression, it is assumed that the catalyst weight does not change significantly during the reaction.

The amounts of gas products and the hydrogen consumption were calculated from the gas analyses and the known volumes of the bombs and gas lines and are expressed as wt. % of DAF coal except for blank runs with solvent only where the basis is wt. % of solvent.

The +525^{0}C residue was determined by TGA. The results are expressed as wt. % of CS_2 solubles.

SIMULATED DISTILLATION

Simulated distillation experiments were carried on selected samples of CS_2 soluble products. A Hewlett-packard 5880A GC with Simulated Distillation system and software was used to determine the boiling range distribution of the samples by ASTM Method D2887-73. A standard sample was run daily for calibration.

PITCH CONTENT BY THERMOGRAVIMETRIC ANALYSIS (TGA)

A Dupont 1090 Thermal Analyzer with TGA option was used. Carbon disulphide extract was transferred to the weighing pan and held under flowing nitrogen at 80^{0}C to (operationally) constant weight. Temperature was then raised at 10^{0}C/min. to 525^{0}C and held for 15 minutes. CS_2 soluble pitch contents are reported as:

$$\frac{\text{Residue}}{80^0\text{C constant wt.}} \times 100 = \underset{\text{(TGA residue)}}{\text{wt \% pitch}}$$

COAL CLEANING

Sized +100 mesh fractions of raw coals were cleaned by laboratory dense medium methods to a specific gravity giving a 70% mass recovery.

Fines (-100 mesh) were cleaned by froth flotation in a 9 ℓ cell using a Denver D-2 machine (1200 rpm, manual air control) using initially 10 wt. % slurries and pine oil as frother/collector. Flotation was continued with incremental 0.1 lb/ton pine oil addition to a barren froth.

RESULTS AND DISCUSSION

COAL AND SOLVENT SAMPLE ANALYSES

Proximate and ultimate analyses of the test coals are presented in Tables 1 and 2. Suppliers data available are presented in Appendix 2.

Solvent ultimate analyses are presented in Table 2. As shown, hydrogen contents of the solvents used were:

Anthracene Oil (AO)	6.21%
Hydrogenated Anthracene Oil (HAO)	6.88%
B-F Recycle Solvent	8.46%

Thus the low severity catalytic hydrogenation treatment had only a slight effect on AO hydrogen content (a crude index of solvent hydrogen donor capability). TGA pitch contents of AO, HAO and BF were 2.1%, 0.5, and 3.5% respectively.

The pitch contents of HAO and BF solvents were determined by replicate analyses, as they were used in calculation of the maximum amounts of coal derived distillate in CS_2 extracts of reaction products (see below, Distillate Yields from Test Coals).

EVALUATION OF PROCESS PARAMETERS

The process variables selected for liquefaction of the three coals in tubing bomb reactors were:

- solvent origin
- solvent/coal ratio
- catalyst type
- temperature
- reaction time

Project time and cost constraints did not allow full factorial experimental design. Instead, selected sets of experiments were performed in a way designed to allow the most important comparisons to be made.

Tables 3, 4 and 5 present all the experimental results of tubing bomb tests on coals and solvents together with run numbers and conditions.

Table 6 presents data on distillate yields from the coals for selected tests.

Tables 7-12 present the results regrouped according to process parameters selected for evaluation.

Table 13 presents simulated distillation data for solvents and selected solvent-coal liquid solutions.

With reference to these tabulated data, the precision of experiments was calculated by analysis of variance of twelve groups of replicates for CS_2 solubles and gas, and of ten groups of replicates for +525°C residue. Standard deviation for CS_2 solubles is 3.6, for gas 0.5 and for +525°C residue, 1.6 of respective units.

DISTILLATE YIELDS FROM TEST COALS

Table 6 presents distillate yield data for test conditions selected as suitable models of the above noted processes.

Yield data are presented as calculated minimum and maximum value for coal derived distillate.

Minimum values were computed with the assumption that all CS_2 soluble pitch was derived from coal (i.e. assumed elimination of solvent pitch during reaction). Maximum values were computed with the assumption that solvent pitch content was not reduced during liquefaction tests.

It is considered that pitch contents of solvent blank run products (Table 5) are not necessarily a reliable guide to levels of solvent derived pitch in liquefaction test products. This is primarily because there is considerable chemical interaction (e.g. H exchange, condensation) between coals and solvents at reaction temperatures. Also, the action of catalysts on solvent alone is considered likely to be different from catalyst action in the presence of both coal solids and coal derived liquids dissolved in solvent.

Although it might be possible to separate the run specific solvent contributions to the +525°C residue, this was not attempted as it would require an extensive investigation of solvent behavior under the liquefaction conditions., The several runs for the solvent blanks were undertaken to provide only a rough indication of the solvent behavior under their conditions.

Inspection of the data of Table 5 on pitch contents of solvent blanks indicates that solvent pitch contents are reduced but not eliminated (i.e. within the range for minimum-maximum coal distillate yields. Anomalous high pitch contents in runs with HAO solvent/HDS catalyst at 460°C and BF solvent/RM catalyst at 480°C could be due to a variety of factors including coking and/or inadvertent inclusion of catalyst in the TGA sample and/or experimental margin of error.

With respect to distillate yield, conversions of the Greenhills and Monkman coals appear to be similar and generally lower than values obtained in parallel tests on Illinois No. 6 coal.

With reference to this and following comparisons, it should be noted that the experimental program was designed to differentiate coal reactivities and identify process(es) having the best potential for economic application to the test coals. Yields from coal liquefaction are dependent on many parameters (time, temperature, pressure, solvent, catalyst age, etc.).

Results of tests under selected conditions can be used to compare coals and/or processes, but specific test data (e.g. liquid yields) should be regarded as a qualitative indication of actual pilot or industrial system performance.

Differences in coal reactivity are least for tests with high levels of red mud catalyst and BF solvent. At 460°C, parallel tests on the three coals resulted in distillate yields as follows:

Coal	Distillate (wt. % DAF coal)	
	Minimum	Maximum
Illinois #6	44.5	52.3
Greenhills	42.9	50.8
Monkman	41.0	49.3

However, distillate yield from the Illinois coal increased to 54.2%/61.9% (minimum/maximum) at 480°C, whereas yields from Greenhills and Monkman were reduced at 480°C in parallel tests.

The maximum distillate yields from Illinois No. 6 coal (62.3% with HAO solvent/HDS catalyst and 61.9% with BF solvent/red mud catalyst) were 11-12% higher than maximum observed yields from the B.C. coals.

With HDS catalyst, or no catalyst, distillate yields from Illinois No. 6 coal were substantially higher than from B.C. coals when compared at either equivalent conditions (e.g. 440°C, high HDS catalyst) or maximum observed yield conditions (e.g. 440°C, high HDS catalyst for Illinois No. 6 compared to all observed yields for Greenhills and Monkman coals).

EVALUATION OF MODELLED PROCESSES

The experimental conditons were designed to (approximately) model reaction conditions in three types of processes which have been used and/or evaluated at large scale:

Bergius Process (Red mud catalyst)
H Coal Process (ebulliated bed of supported Co-Mo Catalyst)
Donor Solvent Processes (uncoupled catalytic hydrogenation of solvent; example: UKNCB, EDS)

It is recognized that the validity of this modelling is limited by a number of factors including:

- lack of optimization data
- non-equilibrium nature of solvents
- inherent differences between batch testing and continuous operation.

However, use of standardized conditions and the (relatively) good precision of most experimental data are considered to mean that comparisons of coal reactivities and solvent/catalyst performances are qualitatively valid.

Published data on performance of Illinois #6 coal in the H-Coal process (1) indicate distillate yields somewhat lower than values obtained in this study. This could be due to use of fresh, finely ground catalyst, complete absence of short circuting in the laboratory batch reactor, and/or other factors. However, since the purpose of this study is to establish qualitative comparisons of coal reactivities and process yields, the indicated difference between lab - and pilot plant results is not considered significant.

In general, results presented for B.C. coals in Tables 3 and 6 indicate relatively low yields of both CS_2 soluble and distillate hydrocarbons in the absence of an effective catalyst. This assessment should be considered preliminary, since no serious attempt was made to optimize the gross or donor hydrogen contents of solvents. However, the lack of effect of increasing solvent:coal ratio (Table 8) suggests that initial donor H content of solvent is not a critically important parameter in relation to either CS_2 soluble or distilate yield.

With particular reference to distillate yield, data of Table 6 suggests that use of an effective catalyst would be essential to generate distillate yields which could even approach an economic yield given the relatively high production cost of B.C. mountain coals.

In considering the disposable iron oxide (red mud or equivalent) catalyst option, it should be noted that sulphur contents of B.C. coals tested are grossly inadequate to provide sulphidization which is widely reported to be necessary to attain optimum catalyst effect. (NB tests were not undertaken to confirm this requirement, and such tests would have required use of a different, low sulphur solvent). Availability of adequate quantities of suitable low cost iron bearing materials at locations in B.C. other than coastal ports is also questionable.

For the HDS catalyst, (or similar material), specific tests on catalyst ageing and process optimization would be required, but supply problems would not be as critical as for a disposable iron catalyst and sulphur contents of coals tested are considered likely to be adequate for maintaining presulphided catalyst in sulphided form. In considering any process based on HDS type catalyst, it must be emphasized that distillate yields in the present test program were low in absolute terms, low relative to Illinois No. 6 distillate yields, and low relative to iron catalyzed runs. In fact, results with HDS 1442A were only marginally better than with BF solvent only.

For either catalyst, the economic viability of distillate production in a single process stage appears questionable in view of low yields relative to maximum yields from Illinois No. 6 coal.

In general, with only qualitative reference to process economics, results of this study suggest that some combination of the following factors may be required to develop an economic scenario for liquefaction of the two B.C. coals tested:

- use of a more active catalyst to enhance distillate yield (either in the liquefaction operation or in upgrading of soluble, non-distillate products).
- development of high value market(s) for non-distillate products.
- access to relatively high price markets (i.e. offshore) for the low sulphur, low ash (soluble) conversion product or (low sulphur, high ash) whole residue.

EFFECT OF SOLVENT

With respect to conversion of coal to CS_2 soluble product, it appears that yields increase (with solvent hydrogen content) in the order:

$$AO < HAO < BF$$

In the absence of added catalyst (Table 7, first data set for each coal) the CS_2 soluble conversion yield increases quite substantially (+17.5% for Greenhills, +9.8% for Monkman, +10.5% for Illinois No. 6) in response to a rather marginal (0.6%) difference in hydrogen content between hydrogenated and raw anthracene oil. The incremental increase in CS_2 soluble yield between HAO and BF solvents (6.88% and 8.46% H respectively) is not so marked (+4.9% for Greenhills).

The difference in CS_2 soluble yields from the solvents appears to be reduced but not eliminated in the presence of red mud catalyst.

The data on pitch contents of CS_2 solubles do not show any clear dependence on solvent characteristics.

SOLVENT:COAL RATIO

As shown in Table 8, results of parallel tests at solvent:coal ratios of 2:1 and 3:1 indicate minor or negligible differences in CS_2 soluble yield. The only exception to this general observation is for the test on Greenhills coal with BF solvent and a high level of red mud catalyst at 480°C where - in a single test - a significantly lower CS_2 soluble yield was obtained at the higher solvent:coal ratio. At present, no explanation (other than the possibility of experimental error) can be advanced to explain this anomaly.

For Monkman coal under the same conditions, a higher conversion was obtained at the 3:1 solvent to coal ratio.

This indicated negligible effect of solvent:coal ratio on liquid yield is surprising, particularly for tests with HAO which has a low hydrogen content. It was expected that a 50% increase in availability of hydrogen donor components of the solvent would increase the conversion significantly.

Hydrogen consumption and gas production values are generally higher at the 3:1 solvent ratio. This is consistent with incremental hydrogen consumption by and gas production from the solvent.

EFFECT OF SELECTED CATALYSTS

Results of selected tests indicating effects of catalysis are presented in Table 9.

For the B.C. coals, tests at 460°C indicate that either HDS or red mud has a substantial positive effect on conversion as measured by CS_2 solubility. CS_2 soluble yields increase by >50% even with low levels of red mud compared to the base case of treatment with raw anthracene oil alone.

As expected, hydrogen consumption values increase with increasing (CS_2 soluble) liquid yield.

For the B-F solvent/red mud catalyst combination, the catalyst provides a notable increase in CS_2 soluble products even at low dosage levels compared to the base case of (high hydrogen content) BF solvent only.

Byproduct gas yields from the B.C. coals (compared to the base case of anthracene oil-coal only) indicate that increasing catalyst dosage reduces or maintains gas production. This effect is also observed for Illinois No. 6 coal (compared to HAO - coal only).

Finally, for high severity (480°C) tests with HDS catalyst, CS_2 soluble yields are reduced to levels comparable to the low values obtained with anthracene oil alone at lower severities. This is considered to be due to extensive coking which was observed (as reactor wall deposits) in these tests.

EFFECTS OF REACTION TEMPERATURE

Selected results presented in Table 10 indicate for the B.C. coals only marginally increased CS_2 soluble yields at 480°C compared to results at 460°C.

There appear to be parallel substantial increases in both gas production and hydrogen consumption in tests at 480°C (compared to lower temperature), with no apparent significant reduction in +525°C residues.

RESIDENCE TIME EFFECTS

Selected test data for B.C. coals indicating the effects of residence time at temperature under otherwise equivalent conditions are presented in Table 2. These data are limited in scope, as test time was not considered a critical variable in experimental planning.

Results indicate that conversion continues at a useful rate up to ~25 minutes at temperature but that increasing residence time to 60 minutes gives at best a marginal increase in conversion with substantially increased gas production/hydrogen consumption values and no significant decrease in +525°C residue.

These limited data are consistent with a previous B.C. Research study on a single seam sample of Greenhills coal. For these tests, optimum "total" conversion (i.e. gas plus material soluble in tetrahydrofuran/pyridine) was found to occur in residence times of 20-30 minutes.

CLEAN COAL TESTS

Results of tests on cleaned B.C. coals are presented in Table 12 with results of parallel tests on raw coal. The amount of data is too small to support any firm conclusions, but inspection of the data leads to the following observations:

- for monkman coal/HAO at 460°C there is an apparent increase in CS_2 solubles for clean coal. Such an increase was not observed for clean Greenhills coal, so experimental verification of the single Monkman clean coal test would be required.

- for Monkman coal/BF solvent/HiRM at 460°C there is a minor increase in CS_2 soluble product for clean coal (vs. Monkman raw coal), whereas Greenhills clean coal gives a lower conversion than Greenhills raw coal under the same conditions. (Again, experimental confirmation would be required.)

- low severity HDS/HAO tests indicate little difference between raw - and clean coal performances.

CONCLUSIONS

1. For the two B.C. coals evaluated, there is no major differentiation in reactivity - as measured by distillate yield - under laboratory best yield conditions.

2. The three general process types can be ranked in increasing order of distillate yield as follows:

Donor Solvent (HAO, non-catalytic)	<	H Coal (HAO, HDS)	<	Bergius (BF solvent, red sand catalyst)

   for (model) process conditions showing best conversions for each of the B.C. coals.

3. The B.C. coals produced significantly lower distillate yields than Illinois #6 reference coal under coal specific laboratory best yield conditions. (Note that best (i.e. maximum distillate yield) conditions for B.C. coals did not correspond to those for Illinois #6 coal.)

4. For the Illinois #6 coal, comparison of present results with published data on pilot tests indicates significantly higher distillate yields in our laboratory tests. Therefore, in the absence of directly comparable data for other process types, results of this study should be used to make only qualitative comparisons of process performance and liquid yields.

5. Under some conditions, the B.C. coals produced large amounts of CS_2 soluble non-distillate material ("pitch") which - for conditions and catalysts tested - was highly resistant to further conversion to distillate. Economic separation/utilization of this pitch component could result in higher overall conversion yields than indicated by

the low (in both relative and absolute terms) distillate yields reported above. For conditions tested, results of this study support the conclusion that supported Co-Mo catalysts (e.g. HDS1442A) and sulphided iron oxide (e.g. red mud) catalysts are not sufficiently active to produce attractive conversion of pitch formed in direct liquefaction of the two coals used.

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TABLE 1
TGA PROXIMATE ANALYSES OF COALS AS TESTED

	As Tested				Moisture Free		
	H_2O	V.M.	F.C.	Ash	V.M.	F.C.	Ash
Greenhills	1.6	23.1	63.0	12.3	23.4	64.1	12.5
Clean Greenhills	1.6	25.8	68.5	4.1	26.2	69.6	4.2
Monkman	1.3	19.8	62.9	16.0	20.1	63.7	16.2
Clean Monkman	0.7	24.1	68.9	6.3	24.2	69.4	6.4
Illinois No. 6	3.0	36.3	50.9	9.8	37.4	52.5	10.1
SBRG #1	23.0	30.5	34.3	12.2	39.6	44.6	15.8
SBRG #2	12.8	34.8	41.1	11.3	40.0	47.1	12.9

TABLE 2

ULTIMATE ANALYSES OF GREENHILLS AND MONKMAN COALS AND OF SOLVENTS USED AS SUPPLIED BY CANADIAN MICROANALYTICAL SERVICE

	Composition (wt. %)				
	C	H	N	O	S
Greenhills	76.52	4.41	1.23	4.92	0.78
Monkman	73.82	3.86	0.95	4.76	0.60
Anthracene Oil	91.07	6.21	0.37	2.48	0.70
Hydrogenated Ant. Oil	90.24	6.88	0.49	1.83	2.51
Bergbau-Forschung Recycle Solvent	88.21	8.46	1.37	2.26	0.26

NOTES TO TABLES 3 - 13

1. Coal charge 10 g (proximate analyses per Table 1) in all tests.

2. Reactor initial cold pressure P_i = 1,000 psig H_2 unless otherwise noted.

3. Catalysts

   HDS 1442A (analysis, Appendix 4)

   High Level ("HiHDS") = 2.24 catalyst/test
   (2% of coal as Mo)

   Low Level ("LoHDS") = 0.224 g catalyst/test
   (0.2% of coal as Mo)

   Red Mud (analysis, Appendix 4)

   High Level ("HiRM") = 2.25 g catalyst/test
   (5% of coal as Fe)

   Low Level ("LoRM") = 0.9 g catalyst/test
   (2% of coal as Fe)

4. Sulphur: for HDS, elemental sulphur added at 0.06 g 5/g catalyst (stoichiometric for sulphidization of Co + Mo.

   for RM, elemental sulphur added at 0.06 g/g catalyst (∿50% of stoichiometric for sulphidization of Fe).

TABLE 3

LIQUEFACTION TEST DATA FOR B.C. COALS

GREENHILLS COAL

Run No.	Solvent	Solvent coal (wt. ratio)	Catalyst	Temp. (°C)	Time (min.)	Conversion Products (wt. % DAF coal)		Hydrogen Consumption (g/100 g DAF coal)	TGA Pitch Content of CS_2 Solubles (wt. %)
						Gas	CS_2 Soluble		
219/5	HAO	2	-	460	25	5.5	41.5	1.2	7.1
220/6	HAO	2	-	460	25	5.1	39.3	1.4	6.4
222/5	HAO	2	-	460	25	4.4	41.9	0.9	6.9
223/5	HAO	2	-	460	25	2.9	42.1	1.1	6.4
227/8	HAO	2	HiHDS	440	25	2.5	54.2	2.8	8.2
228/6	HAO	3	-	460	25	4.3	49.4	0.9	6.6
229/5	HAO	3	-	460	25	5.2	42.0	1.1	5.8
244/3	HAO	3	-	460	25	5.6	44.6	1.2	6.9
230/8	HAO	2	HiHDS	420	25	1.3	44.0	2.4	7.4
232/5	HAO	3	-	420	25	1.5	31.0	0.4	3.3
233/3	HAO	3	-	460	10	2.4	33.7	0.3	6.6
234/7	HAO	2	LoHDS	440	25	2.7	42.6	1.5	7.1
235/5	HAO	3	-	480	25	9.4	45.7	2.0	6.0
237/5	BF	2	LoRM	460	25	6.0	59.1	1.7	9.5
238/8	HAO	2	HiHDS	440	10	1.4	39.1	2.1	4.8
239/5	BF	3	LoRM	460	25	7.4	58.3	1.8	6.9
240/7	HAO	2	HiHDS	420	25	1.4	39.9	2.4	7.4
241/7	BF	2	HiRM	460	25	6.2	67.8	2.1	8.3
246/7	BF	2	HiRM	480	25	11.1	70.2	2.8	9.6
247/7	BF	2	HiRM	440	25	3.4	52.1	1.5	10.5
248/3	HAO	2	HiRM	460	25	5.1	58.0	2.4	9.2

TABLE 3

LIQUEFACTION TEST DATA FOR B.C. COALS(Cont'd.)

GREENHILLS COAL

Run No.	Solvent	Solvent coal (wt. ratio)	Catalyst	Temp. (°C)	Time (min.)	Conversion Products (wt. % DAF coal)		Hydrogen Consumption (g/100 g DAF coal)	TGA Pitch Content of CS_2 Solubles (wt. %)
						Gas	CS_2 Soluble		
248/5	BF	2	-	460	25	6.4	46.9	0.9	7.8
249/8	BF	3	HiRM	480	25	12.1	56.4	2.7	9.5
252/7	BF	2	HiRM	440	60	8.7	52.4	2.1	9.8
252/8	HAO	2	HiHDS	440	60	2.9	59.0	3.4	12.2
200/5	AO	2	-	460	25	4.2	25.5	1.4	4.0
201/7	AO	2	HiHDS	460	25	3.9	60.6	3.1	10.3
203/7	AO	2	HiRM	460	25	2.6	52.1	2.7	6.7
208/7	AO	2	HiHDS	460	25	3.3	56.3	3.5	10.4
210/7 (Pi=1400)	AO	2	LoRM	460	25	3.6	44.1	4.0	7.1
212/7	AO	-	HiHDS	480	25	4.7	31.2	3.6	9.4
215/5	AO	2	-	480	25	5.7	29.7	1.7	7.4
CLEAN COAL									
253/7	BF	2	HiRM	460	25	6.8	52.8	2.2	13.9
254/3	HAO	3	-	460	25	7.2	47.7	1.0	7.6
255/7	HAO	2	HiHDS	420	25	2.2	47.8	2.8	7.4

TABLE 3
LIQUEFACTION TEST DATA FOR B.C. COALS(Cont'd.)

MONKMAN COAL

Run No.	Solvent	Solvent coal (wt. ratio)	Catalyst	Temp. (°C)	Time (min.)	Conversion Products (wt. % DAF coal)		Hydrogen Consumption (g/100 g DAF coal)	TGA Pitch Content of CS_2 Solubles (wt. %)
						Gas	CS_2 Soluble		
223/6	HAO	2	-	460	25	4.3	38.2	0.9	7.9
227/7	HAO	2	HiHDS	440	25	2.2	55.1	2.9	7.9
244/5	HAO	3	-	460	25	5.4	41.8	1.3	6.4
229/6	HAO	3	-	460	25	4.9	42.7	1.0	5.4
230/7	HAO	2	HiHDS	420	25	1.9	40.1	2.5	13.2
232/6	HAO	3	-	420	25	1.3	27.1	0.3	3.3
234/8	HAO	2	LoHDS	440	25	2.6	41.1	1.4	7.0
235/6	HAO	3	-	480	25	9.6	43.4	1.7	5.0
237/6	BF	2	LoRM	460	25	5.5	59.1	1.6	11.3
238/7	HAO	2	HiHDS	440	10	0.9	38.8	2.2	5.5
239/3	BF	3	LoRM	460	25	7.1	60.8	1.8	13.8
240/8	HAO	2	HiHDS	420	25	1.3	36.8	2.4	6.1
241/8	BF	2	HiRM	460	25	6.0	66.3	2.0	8.2
243/7	BF	2	HiRM	480	25	11.0	57.5	2.8	10.3
246/8	BF	2	HiRM	480	25	10.7	63.7	2.8	9.9
247/8	BF	2	HiRM	440	25	3.3	51.4	1.5	7.9
251/7	BF	3	HiRM	480	25	11.8	68.0	2.5	10.7
200/6	AO	2	-	460	25	5.4	30.3	1.1	6.5
201/8	AO	2	HiHDS	460	25	2.8	56.2	3.1	10.3

TABLE 3
LIQUEFACTION TEST DATA FOR B.C. COALS(CONT'd.)

MONKMAN COAL

Run No.	Solvent	Solvent coal (wt. ratio)	Catalyst	Temp. (^{0}C)	Time (min.)	Conversion Products (wt. % DAF coal)		Hydrogen Consumption (g/100 g DAF coal)	TGA Pitch Content of CS_2 Solubles (wt. %)
						Gas	CS_2 Soluble		
203/8	AO	2	HiRM	460	25	3.0	50.1	2.1	6.9
208/8	AO	2	HiHDS	460	25	1.7	57.9	2.9	12.0
210/8 (Pi=1400)	AO	2	LoRM	460	25	3.3	47.4	3.9	7.9
215/6	AO	2	-	480	25	6.3	36.8	2.3	9.1
CLEAN	COAL								
253/8	BF	2	HiRM	460	25	6.1	59.6	2.0	8.9
254/5	HAO	3	-	460	25	3.6	60.9	1.8	5.7
255/8	HAO	2	HiHDS	420	25	1.5	35.7	2.5	6.1

TABLE 4

ILLINOIS #6 SEAM COAL - LIQUEFACTION TEST DATA

Run No.	Solvent	Solvent coal (wt. ratio)	Catalyst	Temp. (°C)	Time (min.)	Conversion Products (wt. % DAF coal)		Hydrogen Consumption (g/100 g DAF coal)	TGA Pitch Content of CS_2 Solubles (wt. %)
						Gas	CS_2 Soluble		
225/5	BF	2	HiRM	460	25	8.7	81.1	3.0	11.8
226/6	HAO	2	-	460	25	7.9	61.7	1.9	4.4
231/5	HAO	2	-	440	25	5.0	54.8	1.6	-
231/6	HAO	3	-	440	25	5.0	50.9	1.6	-
236/7	HAO	2	LoHDS	440	25	4.2	72.6	2.8	5.7
242/5	HAO	2	-	440	25	4.8	57.5	1.7	5.1
242/3	HAO	3	-	440	25	5.0	61.3	1.7	3.8
224/7	HAO	2	HiHDS	440	25	3.7	77.4	3.8	5.3
250/7	BF	2	HiRM	480	25	11.5	79.0	3.3	8.1
199/6	AO	2	-	460	25	4.7	47.4	1.5	4.4
202/5	AO	2	-	460	25	5.3	55.0	2.1	5.9

TABLE 5
SOLVENTS - TUBING BOMB TEST DATA

Run	Solvent	Mass (g)	Catalyst	Temp. (^{o}C)	Time (min.)	Products (wt. % of solvent)		Hydrogen Consumption (g/100 g charged solvent)
						Gas	+525^{o}C residue	
224/8	HAO	20	HiHDS	440	25	0.4	0.46	1.3
225/6	BF	20	HiRM	460	25	0.6	2.5	0.2
226/5	HAO	20	-	460	25	0.3	3.8	0.03
236/8	HAO	20	LoHDS	440	25	0.2	2.4	0.3
250/8	BF	20	HiRM	480	25	1.6	1.5	0.4
251/8	BF	20	HiRM	480	25	1.7	6.2	0.4
214/7	AO	20	HiHDS	440	25	0.7	0.8	1.5

TABLE 6
DISTILLATE YIELDS FROM SELECTED TESTS

COAL	Solvent (solvent:) coal ratio)	Catalyst	Temp. (°C)	Time (min.)	Distillate (wt. % DAF coal)	
					Minimum*	Maximum*
GREENHILLS	HAO (2:1)	HiHDS	440	25	30.4	31.8
	AO (2:1)	HiHDS	460	25	30.3	35.2
	BF (2:1)	HiRM	460	25	42.9	50.8
	BF (2:1)	HiRM	480	25	41.2	49.1
	BF (2:1)	none	460	25	25.1	33.0
MONKMAN	HAO (2:1)	HiHDS	440	25	31.6	32.9
	AO (2:1)	HiHDS	460	25	21.7	26.8
	BF (2:1)	HiRM	460	25	41.0	49.3
	BF (2:1)	HiRM	480	25	33.5	41.7
ILLINOIS #6	HAO (2:1)	HiHDS	440	25	61.1	62.3
	BF (2:1)	HiRM	460	25	44.5	52.3
	BF (2:1)	HiRM	480	25	54.2	61.9

*Minimum values assume no pitch in solvent after treatment with coal; maximum values assume original solvent pitch content in reaction product.

TABLE 7

EFFECT OF SELECTED SOLVENTS ON COAL CONVERSION UNDER EQUIVALENT TEST CONDITIONS

	Conditions	Solvent	Products (wt.% DAF coal)		H_2 Consumption (g/100 g DAF coal)	TGA Residue
			Gas	CS_2Soluble		
GREENHILLS COAL	(460°/25 no catalyst ratio 2)	AO	4.2	25.5	1.4	4.0
		HAO(average of 2)	3.7	42.0	1.0	6.6
		BF	6.4	46.9	0.9	7.8
	(480°/25 no catalyst)	AO(ratio 2)	5.7	29.7	1.7	7.4
		HAO(ratio 3)	9.4	45.7	2.0	6.0
	(460°/25 HiRM ratio 2)	AO	2.6	52.1	2.7	6.7
		HAO	5.1	57.9	2.4	9.2
		BF	6.2	67.8	2.1	8.3
	(460°/25 LoRM)	AO(Pi=1400) (ratio 2)	3.6	44.1	4.0	7.1
		BF(ratio 3)	7.4	58.3	1.8	6.9
MONKMAN COAL	(460°/25 no catalyst ratio 2)	AO	5.4	30.3	1.1	6.5
		HAO(average of 2)	3.8	40.1	0.9	7.7
	(480°/25 no catalyst)	AO(ratio 2)	6.3	36.8	2.3	9.1
		HAO(ratio 3)	9.6	43.4	1.7	5.0
	(460°/25 HiRM ratio 2)	AO	3.0	50.1	2.1	6.9
		BF	6.0	66.3	2.0	8.2
	(460°/25 LoRM)	AO(ratio 2)	3.3	47.4	3.9	7.9
		BF(ratio 3)	5.5	59.1	1.6	11.3
ILLINOIS #6 COAL	(460°/25 no catalyst ratio 2)	AO(average of 2)	5.0	51.2	1.8	5.2
		HAO	7.9	61.7	1.9	4.4

TABLE 8

EFFECT OF SOLVENT:COAL RATIO ON COAL CONVERSION UNDER EQUIVALENT TEST CONDITIONS

	Conditions	Solvent:coal (wt. ratio)	Products (wt.% DAF coal)		H_2 Consumption (g/100 g DAF coal)	TGA Residue
			Gas	CS_2 Soluble		
GREENHILLS COAL	(HAO 460°/25 no catalyst)	2:1(average of 2)	3.7	42.0	1.0	6.6
		3:1(average of 3)				
	(BF 460°/25 HiRM	2:1	11.1	70.2	2.8	9.6
		3:1	12.1	56.4	2.7	9.5
	(BF 460°/25 LoRM)	2:1	6.0	59.1	1.7	9.5
		3:1	7.4	58.3	1.8	6.9
MONKMAN COAL	(HAO 460°/25 no catalyst)	2:1(average of 2)	3.8	40.1	0.9	7.7
		3:1(average of 2)	5.2	42.2	1.1	5.9
	(BF 480°/25 HiRM)	2:1(average of 2)	10.8	60.6	2.4	9.3
		3:1	11.8	68.0	2.5	10.7
	(BF 460°/25 LoRM)	2:1	5.5	59.1	1.6	11.3
		3:1	7.1	60.8	1.8	13.8
ILLINOIS #6 COAL	(HAO 440°/25 no catalyst)	2:1(average of 2)	4.9	56.2	1.7	5.1
		3:1(average of 2)	5.0	56.1	1.7	3.8

TABLE 9

EFFECT OF SELECTED CATALYSTS ON COAL CONVERSION UNDER OTHERWISE EQUIVALENT TEST CONDITIONS

COAL	Conditions			Catalyst	Products (wt. % DAF coal)		H_2 Consumption (g/100 g DAF coal)	TGA Residue
	Solvent	Temp. (°C)	Time (min.)		Gas	CS_2 Soluble		
GREENHILLS	AO	460	25	none	4.2	25.5	1.4	4.0
				HiHDS	3.6	58.4	3.3	10.4
				HiRM	2.6	52.1	2.7	6.7
				LoRM (Pi=1400)	3.6	44.1	4.0	7.1
	AO	480	25	none	5.7	29.7	1.7	7.4
				HiHDS	4.7	31.2	3.6	9.4
	HAO	440	25	LoHDS	2.7	42.6	1.5	7.1
				HiHDS	2.5	54.2	2.8	8.2
	BF	460	25	none	6.4	46.9	0.9	7.8
				LoRM	7.4	58.3	1.8	6.9
				HiRM	6.2	67.8	2.1	8.3
	HAO	420	25	none	1.5	31.0	0.4	3.3
				HiHDS	1.4	41.9	2.4	7.4
MONKMAN	AO	460	25	none	5.4	30.3	1.1	6.5
				HiHDS	2.8	56.2	3.1	10.3
				HiRM	3.0	50.1	2.1	6.9
				LoRM	3.3	47.4	3.9	7.9
	HAO	440	25	LoHDS	2.6	41.1	1.4	7.0
				HiHDS	2.2	55.1	2.9	7.9

TABLE 9 (Cont'd.)

EFFECT OF SELECTED CATALYSTS ON COAL CONVERSION UNDER OTHERWISE EQUIVALENT TEST CONDITIONS

COAL	Conditions			Catalyst	Products (wt. % DAF coal)		H_2 Consumption (g/100 g) DAF coal)	TGA Residue
	Solvent	Temp. (°C)	Time (min.)		Gas	CS_2 Soluble		
MONKMAN (Cont'd.)	BF	460	25	LoRM	5.5	59.1	1.6	11.3
				HiRM	6.0	66.3	2.0	8.2
	HAO	420	25	none	1.3	27.1	0.3	3.3
				HiHDS	1.6	38.0	2.5	9.7
ILLINOIS #6	HAO	440	25	none	4.9	56.2	1.7	5.1
				LoHDS	4.2	72.6	2.8	5.7
				HiHDS	3.7	77.4	3.8	5.3

TABLE 10

EFFECT OF TEMPERATURE ON CONVERSION UNDER OTHERWISE EQUIVALENT TEST CONDITIONS

COAL	Conditions			Temp. (°C)	Products (wt. % DAF coal)		H_2 Consumption (g/100 g) DAF coal)	TGA Residue
	Solvent (Solvent: coal ratio)	Catalyst	Time (min.)		Gas	CS_2 Soluble		
GREENHILLS	AO (2:1)	none	25	460	4.2	25.5	1.4	4.0
	AO (2:1)	none	25	480	5.7	29.7	1.7	7.4
	HAO (3:1)	none	25	420	1.5	31.0	0.4	3.3
*	HAO (3:1)	none	25	460	5.0	45.3	1.1	6.4
	HAO (3:1)	none	25	480	9.4	45.7	2.0	6.0
	HAO (2:1)	HiHDS	25	420	1.4	41.9	2.4	7.4
	HAO (2:1)	HiHDS	25	440	2.5	54.2	2.8	8.2
	B-F (2:1)	HiRM	25	440	3.4	52.1	1.5	10.5
	B-F (2:1)	HiRM	25	460	6.2	67.8	2.1	8.3
	B-F (2:1)	HiRM	25	480	11.1	70.2	2.8	9.6

*Average of 3.

TABLE 10 (Cont'd.)
EFFECT OF TEMPERATURE ON CONVERSION UNDER OTHERWISE EQUIVALENT TEST CONDITIONS

COAL	Conditions			Temp. (°C)	Products (wt. % DAF coal)		H_2 Consumption (g/100 g) DAF coal)	TGA Residue
	Solvent (Solvent: coal ratio)	Catalyst	Time (min.)		Gas	CS_2 Soluble		
MONKMAN	AO (2:1)	none	25	460	5.4	30.3	1.1	6.5
	AO (2:1)	none	25	480	6.3	36.8	2.3	9.1
	HAO (3:1)	none	25	420	1.3	27.1	0.3	3.3
	HAO (3:1)	none	25	460 (avg.of 2)	5.2	42.2	1.1	5.9
	HAO (3:1)	none	25	480	9.6	43.4	1.7	5.0
	HAO (2:1)	HIHDS	25	420 (avg.of 2)	1.6	38.0	2.5	9.7
	HAO (2:1)	HIHDS	25	440	2.2	55.1	2.9	7.9
	B-F (2:1)	HIRM	25	440	3.3	51.4	1.5	7.9
	B-F (2:1)	HIRM	25	460	6.0	66.3	2.1	8.2
	B-F (2:1)	HIRM	25	480	10.9	60.6	2.4	10.1
ILLINOIS #6	HAO (2:1)	none	25	440	4.9	56.2	1.7	5.1
	HAO (2:1)	none	25	460	7.9	61.7	1.9	4.4
	B-F (2:1)	HIRM	25	460	8.7	81.1	3.0	11.8
	B-F (2:1)	HIRM	25	480	11.5	79.0	3.3	8.1

TABLE 11

RESIDENCE TIME EFFECTS

Coal	Conditions	Time (min.)	Conversion Products (wt. % DAF coal)		Hydrogen Consumption (g/100 g DAF coal)	TGA Pitch (wt. % CS_2 solubles)
			Gas	CS_2 Soluble		
GREENHILLS	HAO solvent, 3:1 ration, no catalyst 460°C	10	2.4	33.7	0.3	6.6
		25	5.0*	45.3*	1.1*	6.4*
	HAO solvent, 2:1 ratio, HiHDS catalyst 440°C	10	1.4	39.1	2.1	4.8
		25	2.5	54.2	2.8	8.2
		60	2.9	59.0	3.4	12.2
	BF solvent, 2:1 ratio, HiRM catalyst 440°C	25	3.4	52.1	1.5	10.5
		60	8.7	52.4	2.1	9.8
MONKMAN	HAO solvent, 2:1 ratio, HiHDS catalyst, 440°C	10	0.9	38.8	2.2	5.5
		25	2.2	55.1	2.9	7.9

*Average of 3 runs.

TABLE 12
COMPARISON OF RESULTS WITH RAW & CLEAN COALS

Coal	Test Conditions	Raw Coal			Clean Coal		
		Conversion Products (wt. % DAF coal)		TGA Pitch (wt. % CS_2 solubles)	Conversion Products (wt. % DAF coal)		TGA Pitch (wt. % CS_2 solubles)
		Gas	CS_2 Soluble		Gas	CS_2 Soluble	
GREENHILLS	BF solvent, 2:1 ratio, HiRM, 460°C, 25 min.	6.2	67.8	8.3	6.8	52.8	13.9
	HAO solvent, 3:1 ratio, no catalyst, 460°C, 25 min.	4.3 5.2	49.4 42.0	6.6 5.8	7.2	47.7	7.6
	HAO solvent, 2:1 ratio HiHDS, 420°C, 25 min.	1.3	44.0	7.4	2.2	47.8	7.4
MONKMAN	BF solvent, 2:1 ratio, HiRM, 460°C, 25 min.	6.1	59.6	8.9	6.0	66.3	8.2
	HAO solvent, 3:1 ratio, no catalyst, 460°C, 25 min.	5.4 4.9	41.8 42.7	6.4 5.4	3.6	60.9	5.7
	HAO solvent, 2:1 ratio, HiHDS, 420°C, 25 min.	2.4	36.8	6.1	1.5	35.7	6.1

TABLE 13

SIMULATED DISTILLATION TEST DATA ON RAW AND HYDROTREATED SOLVENTS, AND SOLVENT-COAL LIQUID SOLUTIONS

Raw Material	Run No.	Temp. (°C)	Time (min.)	Catalyst	Distillate Distribution (Volume %) IBP 200°C	Distillate Distribution (Volume %) IBP 200-325°C	Distillate Distribution (Volume %) FBP 325°C	Final Boiling Point (°C)
Raw Solvents								
Anthracene Oil					3	68	29	428
Hydrogenated Anthracene Oil					3	66	31	444
Bergbau Forschung Solvent					4	46	50	498
Hydrogenated Solvents								
Anthracene Oil	214/7	440	25	Red Mud	4	75	21	381
	214/8	440	25	HDS	3	71	26	414
AO-Greenhills (2:1)	200/5	460	25	none	2	70	28	488
- do. -	201/7	460	25	HiHDS	5	69	26	487
- do. -	208/7	460	25	HiHDS	9	68	23	409
- do. -	203/7	460	25	HiRM	5	71	24	392
- do. -	210/7	460	25	LoRM	3	71	26	393
- do. -	215/5	460	25	none	3	72	25	392
HAO-Greenhills (2:1)	223/5	460	60	none	3	68	29	445
- do. -	252/8	440	25	HiHDS	4	48	48	479
BF-Greenhills (2:1)	246/7	480	25	HiRM	9	52	39	484
- do. -	252/7	440	60	HiRM	4	38	58	500
AO-Monkman (2:1)	200/6	460	25	none	1	71	28	461
- do. -	201/8	460	25	HiHDS	4	72	24	449

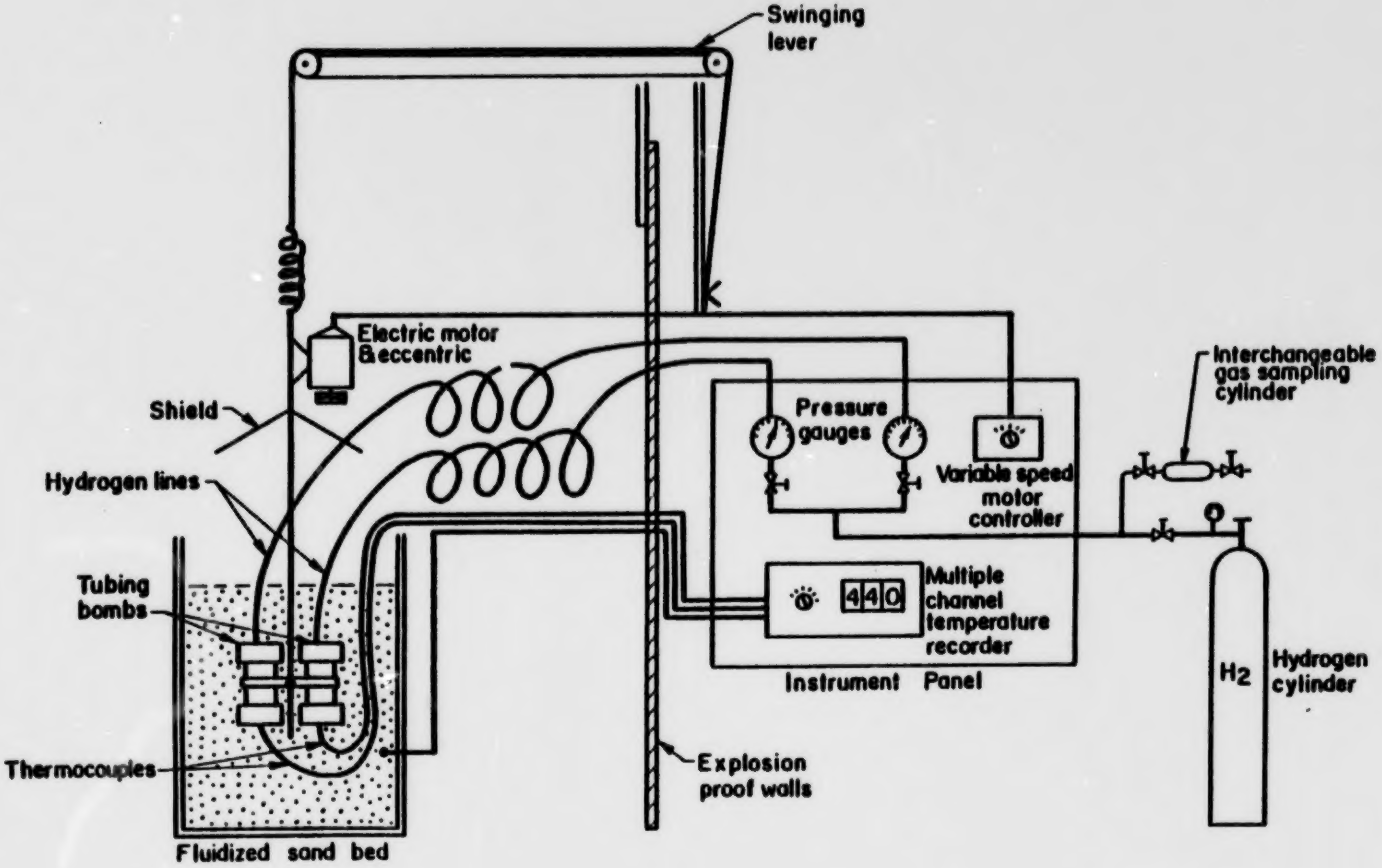
TABLE 13 (Cont'd.)

SIMULATED DISTILLATION TEST DATA
ON RAW AND HYDROTREATED SOLVENTS, AND SOLVENT-COAL LIQUID SOLUTIONS

					Distillate Distribution (Volume %)			Final Boiling Point (°C)
					IBP 200°C	IBP 200-325°C	FBP 325°C	
Raw Solvents								
Anthracene Oil					3	68	29	428
Hydrogenated Anthracene Oil					3	66	31	444
Bergbau Forschung Solvent					4	46	50	498
Hydrogenated Solvents								
Raw Material	Run No.	Temp. (°C)	Time (min.)	Catalyst				
Anthracene Oil	214/7	440	25	Red Mud	4	75	21	381
	214/8	440	25	HDS	3	71	26	414
AO-Monkman (2:1)	208/8	460	25	HiHDS	5	71	24	415
- do. -	203/8	460	25	HiRM	9	66	25	420
- do. -	210/8	460	25	LoRM	3	71	26	395
- do. -	215/6	480	25	none	3	73	24	391
HAO-Monkman (2:1)	223/16	460	25	none	3	68	29	490
AO-Illinois #6 (2:1)	199/6	460	25	none	4	65	31	451
- do. -	202/5	460	25	none	0	74	26	389
HAO-Illinois #6 (2:1)	226/6	460	25	none	5	65	30	453
BF-Illinois #6 (2:1)	225/5	460	25	HiRM	7	49	44	471
- do. -	250/7	480	25	HiRM	10	51	39	483

FIGURE 1

SCHEMATIC DIAGRAM OF TUBING BOMB REACTOR


Swinging lever
Electric motor & eccentric
Shield
Hydrogen lines
Tubing bombs
Thermocouples
Fluidized sand bed
Explosion proof walls
Pressure gauges
Variable speed motor controller
440
Multiple channel temperature recorder
Instrument Panel
Interchangeable gas sampling cylinder
H2
Hydrogen cylinder

APPENDIX 1

REVISED TERMS OF REFERENCE

KILBORN ENGINEERING (B.C.) LIMITED

MINUTES OF MEETING NO. 9

Conference Call - September 9, 1981

DATE OF MEETING: Conference Call held on September 9, 1981

PROJECT NO.: 3349-15

PARTICIPANTS:
R. McElroy - B.C. Research, Vancouver
J. Jezko - " "

L. Gutierrez - Petro-Canada, Calgary

G. Teodosiu - SBRG, Toronto
E. Chornet - SBRG, Sherbrooke

A. Shaikh - Kilborn, Toronto

PREPARED BY: A. Shaikh

The conference call was held to discuss the proposed test scheme prepared by R. McElroy and J. Jezko (attached) as discussed in the meeting of September 4, 1981.

1. A. Shaikh enquired whether the new scheme was to override the previous data collected. R. McElroy informed that there is substantial overlap between the previous and new program. Some of the results previously obtained would still be valid.

2. A. Shaikh enquired about the time frame required to perform the tests. R. McElroy assured that barring any technical problems, BCR should be able to complete the tests and have the results circulated in a draft form by the end of September so that SBRG can proceed with their test program.

3. E. Chornet questioned the hydrogenation procedure adopted by BCR. He was satisfied that the procedure, although not a commonly used one, developed by BCR was adequate.

4. L. Gutierrez enquired why BCR was proposing to hydrogenate the anthracene oil in the tubing bomb instead of the autoclave. R. McElroy stated that he had done some calculations on the time involved and the tubing bomb was more economical. To hydrogenate the solvent in the autoclave would cost around $2,000 as against $1,200 in the tubing bomb. However, E. Chornet expressed preference for the hydrogenation in a larger autoclave and conforming to more defined procedures.

cont'd.../2

5. L. Gutierrez enquired the level of the hydrogen content in the solvent. It was decided that BCR would prepare several (about 20) batches of hydrogenated anthracene oil in tubing bombs, keeping the hydrogen level between 7 to 8%. The different batches will be combined to form a composite mixture (about 1 litre and 7.5% total hydrogen content) and characterised. The total quantity (approx. 1 litre) will be adequate for the test runs leaving sufficient quantity for any repeat experiments.

6. G. Teodosiu enquired the method of characterization to be employed by BCR. It was agreed that the composite mixture will be topped to 200°C, do a simulated distillation on it and send it out for elemental analysis.

7. E. Chornet suggested that the 3:1 solvent to coal ratio should be used.

8. G. Teodosiu had reservations about operating the prehydrogenated solvent and catalyst runs at 480°C. It was agreed that for non-catalytic runs 440 and 460°C is acceptable. It was agreed that all tests be performed first at 460°C and depending on the results, the next operating temperature would be decided.

9. G. Teodosiu enquired whether 10% of red mud was in effect 10% of Fe content. This meant that the red mud content was 50% of the coal. It was agreed that the Fe content should be 0.4% for the low level and 5% Fe by wt of DAF coal for the high level.

10. L. Gutierrez pointed out that in the matrix no provision is made for tests with Illinois #6 with hydrogenated anthracene oil and low level HDS catalyst. R. McElroy suggested that this could be used as a spare run.

11. L. Gutierrez reiterated the economic aspect of the test program, which now has 54 runs and also the additional cost of hydrogenating the anthracene oil. A. Shaikh informed that this aspect of the test program will be addressed to the Consortium at the next Technical Committee meeting. There also is a suggestion that the costs could be reduced by not performing simulated distillation on 16 runs, but just doing TGA's. The 16 run on which the simulated distillation is not to be performed will be decided later. (At a conference call to be arranged at a later date.)

12. G. Teodosiu wished to know if the coal analysis determined by BCR could also be provided to SBRG. R. McElroy agreed to send the coal analysis before SBRG starts their program.

cont'd.../3

13. G. Teodosiu wanted to know if red mud was the likely catalyst to be used, so that she could arrange for drying of the red mud slurry.

14. The next conference call will be set for a suitable date depending on the progress made by BCR.

15. A. Shaikh requested L. Gutierrez that M. Grandbois (absent in this meeting) be informed of the discussion held. If M. Grandbois had any concerns, he should contact A. Shaikh at the earliest.

AMS/wn

PROVISIONAL TEST SEQUENCE PROPOSED BY BCR

PROJECT NO.: 3349-15

REF NO.	COAL	SOLVENT	CATALYST	TIME	TEMP	NO. OF SINGLE BOMB TESTS
1	None	HAO	None	25	460,480	2
2	G,M	HAO	None	25	440,460,480	6
3	16	HAO	None	25	440,460	4
	(With 16, Run G or M coal at 3:1 solvent to coal ratio)					
4	G,M	HAO	HDS	25	440,460	4
5	16	HAO	HDS	25	440,460	2
6	None	HAO	HDS	25	440,460	2
7	G,M	HAO	Low Level HDS	25	Select Temp. (expect 460)	2
8	None	B-F	Red Mud	25	460,480	2
9	16	B-F	Red Mud	25	460,480	2
10	G,M	B-F	Red Mud	25	460,480	4
11	G,M,16 and blank	B-F	Low Level Red Mud	25	Select Temp. (expect 480)	4
12	G,M	B-F (3:1)	Red Mud	25	480	2
13	CG,CM	HAO (Select Ratio)	None	25	Select Temp. (expect 460)	2
14	CG, CM	HAO	HDS	25	Select Temp.	2
			(Select level)	25	Select Temp. (expect 460)	2
15	CG,CM	B-F (Select Ratio)	Red Mud	25	Select Temp. (expect 480)	2

cont'd.../2

Provisional Test Sequence
Proposed by BCR
Project No. 3349-15

REF NO.	COAL	SOLVENT	CATALYST	TIME	TEMP	NO. OF SINGLE BOMB TESTS
16	G,M,16	HAO (Select Ratio)	None	10	Select Temp. (expect 460)	3
17	G,M,16	HAO	HDS	10	Select Temp. (expect 460)	3
18	Spares	---------------------Select-------------------------------				4
						54

NOTE:

ABBREVIATIONS: G = Greenhills; M = Monkman; HAO = Hydrogenated anthracene oil;
16 = Illinois No. 6; B-F = Bergbau-Forschung;
HDS = Sulphided H-Coal Catalyst; CG = Cleaned Greenhills;
CM = Cleaned Monkman.

APPENDIX 2

DATA ON GREENHILLS AND MONKMAN COALS

BC Coal

B.C. Coal Ltd
[illegible] West Georgia Street
Vancouver British Columbia
Canada V6E [illegible]

[illegible]

September 21, 1981

Kilborn Engineering Ltd., 2200 Lakeshore Blvd. W., Toronto, Ontario. M8V 1A4	B. C. Research, 3650 Wesbrook Mall, Vancouver, B.C. V6S 2L2
Attention: Dr. A. Shaikh	Attention: Dr. R. McElroy
Dear Dr. Shaikh:	Dear Dr. McElroy:

Please find enclosed a copy of the results of the samples taken from the Greenhills adits in the proportion of:

No. 10 Seam	25%
7 Seam	25%
1 Seam	50%

The samples were analysed in the laboratory of B. C. Coal.

Yours truly,

B. C. COAL LTD.

D. M. Parkes.

DAVID M. PARKES, P. Eng.,
Director,
Underground Mining Technology.

c.c. Mr. E. Jamieson, Chairman, Liquifaction Technical Committee,
Petrocanada.

Encl.

RESULTS OF B.C. RESEARCH SAMPLE

Liquifaction Study - Greenhills Sample

PROXIMATE ANALYSIS:

Ash %	V.M.	F.C.	F.S.I.	Sul %	B.T.U.	Air dry M% = 4.5%
13.6	23.2	63.2	6½	0.36	14,140/ 7,856	Dry basis

PLASTICITY:

131.6 D.D.M.

SIZE ANALYSIS:

	WT%	Cum WT%
+ ½"	17.3	17.3
-½" x 28m	46.0	63.3
-28 x 35m	7.7	71.0
-35 x 48m	6.5	77.5
-48 x 60m	2.3	79.8
-60 x 100m	6.2	86.0
-100 x 0	14.0	100.0

ULTIMATE:

% Carbon	74.32
% Hydrogen	4.16
% Nitrogen	0.99
% Sulphur	0.38
% Ash	13.67
% Oxygen	6.17
% I.M.	0.31

HARDGROVE GRINDABILITY: INDEX - 107

Ash Fusion Temperatures:	Reducing	Oxidizing
Initial deformation:	2430	2640
Softening:	2600	2710
Hemispherical:	2650	2730
Fluid:	2700	2760

PETROGRAPHY: As per attached Sheet

R_o = 1.12

SPARWOOD B.C.

SAMPLE: BCR' AUG. 20/81 DATE: SEPT. 8 19 81

PROXIMATE ANALYSIS (dry basis)

	raw	
ash %	13.6	
volatile matter %	23.2	
fixed carbon %	63.2	
free swell index	6½	
sulfur %	0.36	
P_2O_5 %		
BTU's/lb. (gross)	7456 14,140	

GIESLER PLASTICITY

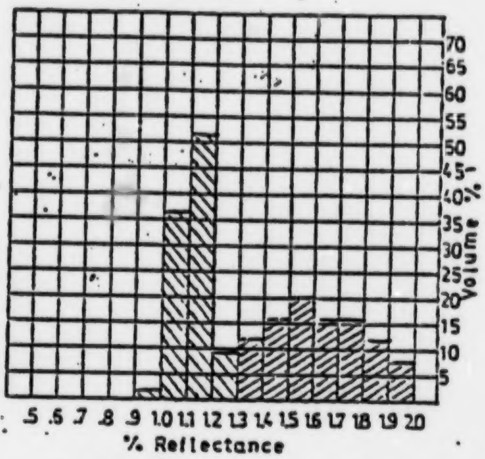
start	403	°c
fusion temp.	428	°c
max. fluid temp.	457	°c
final fluid temp.	480	°c
solidific. temp.	487	°c
melting range	52	°c
max. fluidity	131.6	dd/m

DILATATION (Ruhr)

softening temp.		°c
max. contraction temp.		°c
max. dilatation temp.		°c
contraction		%
dilatation		%
G factor		

PETROGRAPHIC DATA #1459

R_o Type	vitrinoids %	ACT. VOL.	SEMI-FUS. %	ACT. VOL.	Total ACT. VOL.
4					
5					
6					
7					
8					
9	2	1.1			1.1
10	37	20.7			20.7
11	52	29.1			29.1
12	9	5.0			5.0
13			12	1.8	1.8
14			16	2.4	2.4
15			20	3.0	3.0
16			16	2.5	2.5
17			16	2.4	2.4
18			12	1.8	1.8
19			8	1.2	1.2
20					
Vol.		55.9		15.1	71.0



OXIDATION: NOT NOTICABLE

maceral Composition analysis

vitrinoids (colinite & telinite)	%	55.9
reactive semi-fusinoids	%	15.1
exinoids (spor., res., cut.)	%	0.3
total reactive components	%	71.3
inert semi-fusinoids	%	15.1
macrinoids & micrinoid	%	4.0
fusinoids (pyro & degrado) & sclerotinite	%	1.8
mineral matter	%	7.8
other	%	
total inert components	%	28.7

mean max. reflectance ($\bar{R}_o$) of vitrinite 1.12
$\bar{R}_o$ of semi-fusinite 1.62
R_o overlap NONE
balance index 1.07 strength index 4.26
predicted stability (petrographic) 58.3

ASH MINERAL ANALYSIS

SiO_2 - 59.78

Al_2O_3 - 28.03

TiO_2 - 1.62

Fe_2O_3 - 4.15

CaO - 1.68

MgO - 0.88

Na_2O - 0.35

K_2O - 0.90

P_2O_5 - 0.97

SO_3 - 0.62

Undetermined 1.02

BCR

[redacted]

Energy Mines and Resources Canada
Énergie, Mines et Ressources Canada

CANMET
Canada Centre for Mineral and Energy Technology
Centre canadien de la technologie des minéraux et de l'énergie

A STUDY OF THE CAKING AND COKING CHARACTERISTICS OF CLEANED ADIT SAMPLES FROM SEAMS B1, B4, B5 AND B9 FROM THE MONKMAN COAL PROJECT IN N.E. BRITISH COLUMBIA SUBMITTED BY PETRO-CANADA

Project 03-3-1/29-2
Job No. 3227R

J.G. JORGENSEN, W. GARDINER, AND T.A. LLOYD
COAL RESOURCE AND PROCESSING LABORATORY

MARCH 1980

ENERGY RESEARCH PROGRAM
ENERGY RESEARCH LABORATORIES
REPORT ERP/ERL 80-34 (CF)

~~CONFIDENTIAL~~
Declassification Date:
April 1981

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(Project 03-3-1/29-2)
Job No. 3227R

by

J.G. Jorgensen*, W. Gardiner*, and T.A. Lloyd*

INTRODUCTION

This investigation deals with the evaluation of four cleaned adit coal samples from seams B1, B4, B5 and B9 from the Monkman Coal Project in Northeastern British Columbia. The project was initiated by J.Y. Wright, Chief Geologist, Petro-Canada, in a letter dated June 6, 1979, which is reproduced in Appendix 1.

The raw adit samples were cleaned by Birtley Engineering (Canada) Ltd., Calgary. The bulk samples were carbonized in the 12-inch width movable-wall coke oven located at the CANMET Bells Corners Complex near Ottawa. Representative samples were taken and analysed for complete chemical, physical, thermal rheological and petrographical properties. The results of the testing and analyses are listed in Tables 1 to 6.

---

*Heads, Petrography Section, Carbonization Operations Section, and Coal Treatment and Rheology Section, respectively, Coal Resource and Processing Laboratory, Energy Research Laboratories, CANMET, Department of Energy, Mines and Resources, Ottawa, Canada.

TABLE 1 Chemical Analyses of Component Coals

Identification				
Laboratory Number	3805-79	3806-79	3852-79	4120-79
Description	79-B-4	79-B-1	79-B-5	79-B-9
Classification				
Rank (ASTM)	mvb	mvb	mvb	mvb
International System	432	433	432	433
Specific Volatile Index	195	199	191	184
Carbon (dmmfb) ....%	92.1	90.5	89.7	89.2
Proximate Analysis (db)				
Ash ....%	7.1	6.3	8.7	7.7
Volatile Matter ....%	22.4	22.3	22.7	25.9
Fixed Carbon ....%	70.5	71.4	68.6	66.4
Gross Calorific Value (db)				
Btu per pound	14455	14645	14105	14240
Ultimate Analysis (db)				
Carbon ....%	84.9	84.2	81.0	81.6
Hydrogen ....%	5.0	4.7	4.4	4.7
Sulphur ....%	0.26	0.79	0.55	0.51
Nitrogen ....%	1.0	1.0	1.1	0.8
Ash ....%	7.1	6.3	8.7	7.7
Oxygen (by difference) ......%	1.7	3.0	4.2	4.7
Ash Analysis (db)				
SiO_2 ....%	53.9	57.4	49.7	56.2
Al_2O_3 ....%	23.3	25.6	26.3	25.6
Fe_2O_3 ....%	4.6	7.1	5.3	5.6
TiO_2 ....%	1.4	1.0	1.2	1.2
P_2O_5 ....%	1.5	0.2	4.1	2.3
CaO ....%	4.0	1.0	4.4	2.7
MgO ....%	1.4	0.4	0.9	0.8
SO_3 ....%	4.7	1.3	1.9	0.9
Na_2O ....%	1.1	0.5	0.5	0.3
K_2O ....%	0.5	1.4	1.0	1.3

TABLE 2 Physical Tests and Fusibility of Ash of Component Coals

Identification					
Laboratory Number		3805-79	3806-79	3852-79	4120-79
Description		79-B-4	79-B-1	79-B-5	79-B-9
Coal Pulverization					
Sieve Analysis					
Passing	Retained On				
	1/4 in. %	0.2	0.1	0.1	0.1
1/4 in.	1/8 in. %	12.1	11.8	10.0	10.8
1/8 in.	1/16 in. %	17.2	18.3	15.6	18.2
1/16 in.	1/32 in. %	16.7	19.2	17.7	20.8
1/32 in.	..........%	53.8	50.6	56.6	50.1
Total Passing	1/8 in. %	87.7	88.1	89.9	89.1
Grindability					
Hardgrove Index		85	90	86	77
Fusibility of Ash					
Initial Deformation Temp. ... °F		2160	2380	2240	2570
Softening Temp. Spherical ... °F		2620	2700+	2600	2700+
Softening Temp. Hemispherical °F		2700+	2700+	2700+	2700+
Fluid Temp. .... °F		2700+	2700+	2700+	2700+

TABLE 3 Thermal Rheological Properties of Component Coals

Identification				
Laboratory Number	3805-79	3806-79	3852-79	4120-79
Description	79-B-4	Hunkman 79-B-1	79-B-5	79-B-9
Linear Expansion				
Bd. 52 lb/ft^3 at 2% moisture...%	-12.5	-7.3	-14.7	-17.8
Gieseler Plasticity				
Start ....°C	439	429	432	428
Fusion Temp. ....°C	462	444	447	440
Max. Fluid Temp. ....°C	466	466	465	460
Final Fluid Temp. ....°C	482	491	487	481
Solidification Temp. ....°C	489	495	492	485
Melting Range ....°C	43	62	55	53
Max. Fluidity ....dd/m	8.1	117.5	35	49.2
Torque ....g.in.	40	40	40	40
Dilatation				
TI - Softening Temp. ......°C	416	394	400	395
TII - Max. Contraction Temp. °C	455	442	450	444
TIII - Max. Dilatation Temp. °C	474	470	479	468
Contraction ....%	24	26	26	28
Dilatation ....%	-16	35	-9	0
Free Swelling Index				
F.S.I.	6½	8	7	7

TABLE 4 Petrographic Analysis of Component Coals

Identification				
Laboratory Number	3805-79	3806-79	3852-79	4120-79
Description	79-B-4	79-B-1	79-B-5	79-B-9
Distribution of Vitrinite Types				
V-6 %				
V-7 %				
V-8 %				
V-9 %				
V-10 %			1.1	10.9
V-11 %	4.7	1.8	4.6	33.9
V-12 %	32.2	17.0	40.1	9.9
V-13 %	15.1	37.7	10.9	
V-14 %		4.3	0.6	
V-15 %				
V-16 %				
V-17 %				
V-18 %				
Reactive Components				
Total Vitrinite %	52.0	60.8	57.3	54.7
Reactive Semi-fusinite (1/2) %	14.6	12.5	12.6	13.3
Exinite %	0.0	0.0	0.0	0.6
Total %	66.6	73.3	69.9	68.6
Inert Components				
Inert Semi-fusinite (1/2) %	14.6	12.6	12.7	13.4
Micrinite %	6.9	6.8	6.4	4.6
Fusinite %	8.0	3.7	6.1	9.1
Mineral Matter %	3.9	3.6	4.9	4.3
Total %	33.4	26.7	30.1	31.4
Petrographic Indices				
Mean Reflectance %	1.26	1.31	1.25	1.15
Balance Index	2.15	1.80	1.80	1.47
Strength Index	4.94	5.60	4.90	4.34
Stability Index	52.4	59.5	54.5	53.4

TABLE 5 - Carbonization Data

Test Identification Number	746		742		758		747	
Data of Test	10 Dec./79		15 Nov./79		8 Feb./80		11 Dec./79	
Laboratory Number	3805-79		3806-79		3852-79		4120-79	
Description	79-B-4		79-B-1		79-B-5		79-B-9	
CARBONIZATION DATA								
Net Weight of Charge (wet) lb	567.5		538.1		554.8		551.8	
Moisture in Charge %	3.0		3.0		2.8		3.0	
ASTM Bulk Density (wet) lb/ft^3	-		-		-		-	
Oven Bulk Density (db) lb/ft^3	51.8		49.1		50.7		50.3	
CARBONIZATION RESULTS								
Gross Coking Time hr:min	9:20		8:45		8:45		9:05	
Maximum Wall Pressure lb/in^2	0.7		1.2		0.3		0.4	
Coke Yield Actual %	77.9		77.9		77.5		75.2	
Mean Coke size in	1.86		1.96		1.94		1.89	
Apparent Specific Gravity	0.99		0.93		0.95		0.96	
Screen Analysis of Coke (cumulative percentage retained on)								
3 inch sieve	5.8		7.1		6.6		4.8	
2 inch sieve	36.6		40.7		42.9		39.9	
1 1/2 inch sieve	67.4		72.4		71.4		69.3	
1 inch sieve	92.7		93.9		92.0		92.8	
3/4 inch sieve	94.9		96.3		94.3		95.3	
1/2 inch sieve	95.7		97.2		95.2		96.4	
Percentage -1/2 inch (breeze)	4.3		2.8		4.8		3.6	
Tumbler Test (ASTM)								
Stability Factor	52.7		59.5		52.1		45.5	
Hardness Factor	65.1		71.5		64.7		63.9	
Japanese Drum Test (JIS) (cumulative percentage retained on)	*	**	*	**	*	**	*	**
50 mm sieve	9.1	1.6	19.9	4.0	17.3	2.0	15.3	-
25 mm sieve	86.4	68.0	86.2	71.3	86.6	70.5	85.1	66.2
15 mm sieve	92.6	81.2	96.6	83.2	93.2	81.0	91.8	77.5

*30 revs **150 revs

TABLE 6

Analyses of Coke Oven Charges and Resultant Cokes

Identification				
Test Number.............	746	742	758	747
Date Charged............	10/12/79	15/11/79	8/2/80	11/12/79
Description.............	79-B-4	79-B-1	79-B-5	79-B-9
Coke Oven Charge				
Laboratory Number.........	3805-79	3806-79	3852-79	4120-79
Proximate Analysis (db)				
Ash....................%	7.1	6.3	8.7	7.7
Volatile Matter........%	22.4	22.3	22.7	25.9
Fixed Carbon ....%	70.5	71.4	68.6	66.4
Sulphur (db).............%	0.26	0.79	0.55	0.51
Resultant Coke				
Laboratory Number.........	2032-80	2028-80	2034-80	2033-80
Proximate Analysis (db)				
Ash....................%	8.7	8.0	11.1	9.9
Volatile Matter........%	2.0	1.8	1.3	1.8
Fixed Carbon...........%	89.3	90.2	87.6	88.3
Sulphur (db).............%	0.20	0.70	0.52	0.45

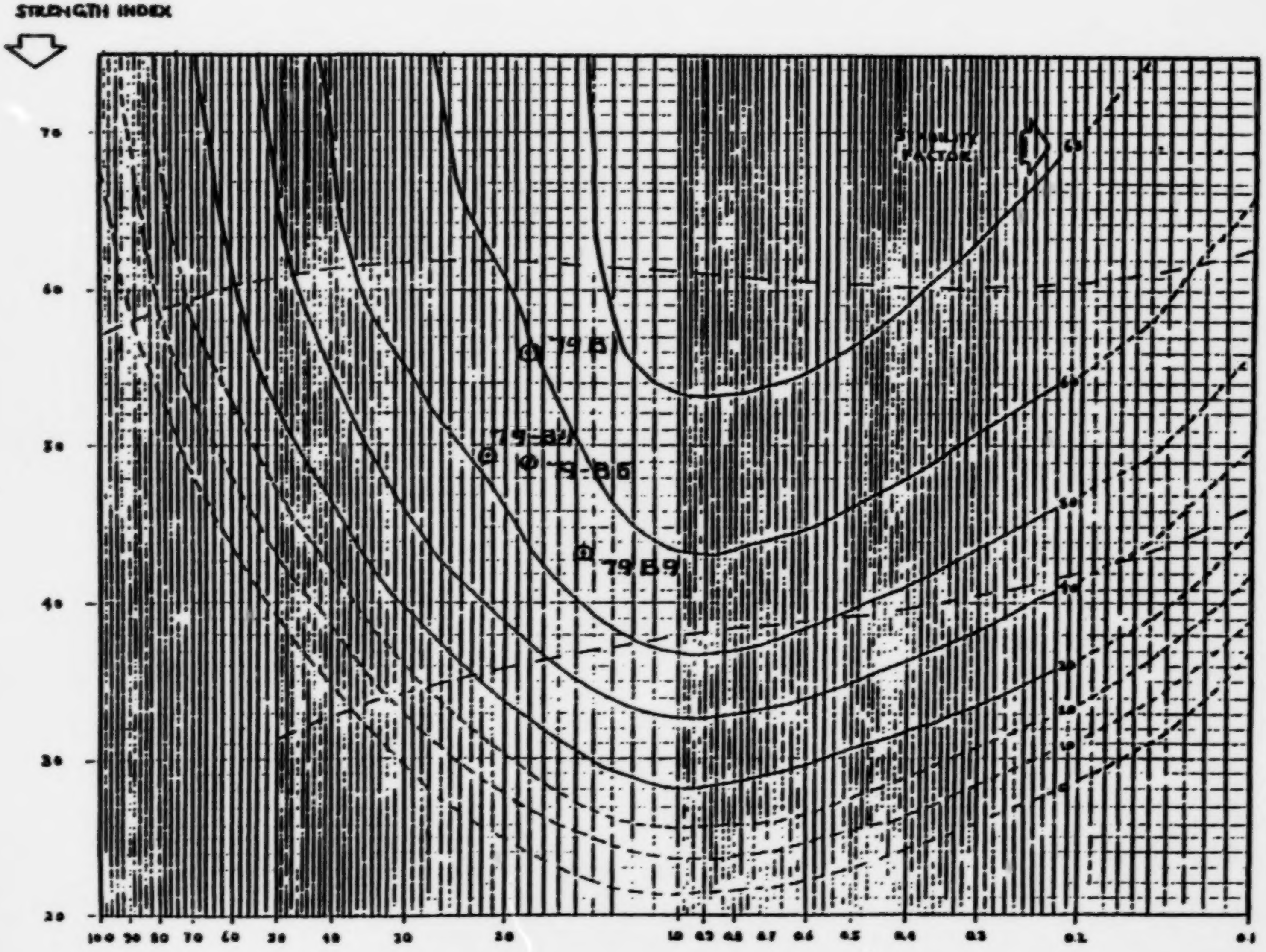

Figure 1. Plot of Predicted Stability Factors from Petrographic Data for Component Coals.

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APPENDIX 3

CATALYST ANALYSES

Catalysts Compositions

American Cyanamid HDS-1442A:

Mo	8.92	wt. %
Co	2.15	wt. %
Ti	0.13	wt. %
B	0.02	wt. %
Fe	0.007	wt. %
Al_2O_3	86.33	wt. %

Red Mud catalyst:

Fe (as Fe_2O_3)	22.1	wt. %

SBRG Continuous Tests

PREFACE

Sandwell Beak Research Group was retained by Kilborn Engineering to conduct a testing program to investigate yields and product quality from the direct liquefaction of two coals from British Columbia in SBRG's continuous liquefaction process development unit.

The principal investigators for this project were

Mr. S.L. Hodd, Executive Director
Dr. E. Chornet, Scientific Advisor
Mrs. G. Teodosiu, Project Manager
Mr. R.H. Arsenault, Testing Supervisor
Mrs. R. Lega, Chemist
Dr. C.A. Johnson, Scientific Advisor

The KILBORN project number was 3349-15
The SBRG project number was R5094

SUMMARY

The overall objective of this project was the comparative investigation of yields and product quality from the direct liquefaction of Monkman and Greenhills coals in SBRG's continuous liquefaction process development unit. Each coal was tested in once-through mode of operation at one set of conditions using a different catalyst for each coal. The catalysts and the experimental parameters were selected by the client based on the analysis of the results obtained in batch testing of the coals by B.C. Research.

The program also included the generation of a hydrogenated anthracene oil to be used as solvent in the continuous tests instead of the coal derived solvent initially proposed due to unavailability of the latter.

The two coals supplied, ground by the client, were tested at the selected conditions without complete shut-down of the unit; mass balances, conversion, gas, liquid and distillate yields were calculated for each test.

About 190 Kg of hydrogenated solvent were generated using a Co-Mo catalyst. This solvent was subsequently used in the two coal liquefaction tests described below.

Experiment 13
Monkman coal & hydrogenated solvent & HDS1442 Co-Mo catalyst & sulphur & H_2
$T = 440^{\circ}C$, $P = 20.7$ MPa

Experiment 14
Greenhills coal & hydrogenated solvent & red mud catalyst & sulphur & H_2
$T = 440^{\circ}C$, $P = 26.6$ MPa

Approximately 40 Kg of product were generated in experiment 13 and 30 Kg in experiment 14.

At the conditions of the tests the conversions obtained (basis: THF insoluble m.a.f. at 25°C) were 82.8% for Monkman coal with Co-Mo catalyst and 76.6% for the Greenhills coal with red mud catalyst; the liquid yields (defined as THF solubles at 25°C) 68.2% and 65.6% respectively and the distillable yield (<480°C) in the order of min. 19%/max 34% for Monkman and min. 17.7/max 33.8% for Greenhills.

In addition to the liquid yields obtained, the yields of distillable product appear low. These relatively low results agree with the results obtained by other groups (see EXXON reports in Appendix A) when working with coals having similarly low volatile matter content.

In view of the low distillate yields obtained at the conditions tested, the probability of these coals being economically viable feedstocks for one step direct liquefaction appears remote unless:

- pretreatment of these coals can lead to physico-chemical and structural changes enhancing their reactivity.
- drastically different reaction conditions succeed in breaking down some of the inert maceral structures and limit the extent of repolymerization of the solubilized fractions. This comment stems from the relatively high liquid yields observed while the distillate yields appear rather low.
- Severe catalytic hydrotreatment of the high molecular weight liquids can lead to improved distillate yield. This implies a two-step liquefaction process.

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EDS Coal Liquefaction Process Development
Annual Technical Progress Report
July 1979 - June 1980
July 1978 - June 1979

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TESTING PROGRAM

1.0 DESCRIPTION OF CONTINUOUS FLOW LIQUEFACTION SYSTEM

The testing program was conducted in Energy Research Laboratories of SBRG using the direct liquefaction Process Development Unit (PDU). The unit was designed for a nominal capacity of 10 kg coal slurry/hour, maximum operating temperature of 475°C, maximum operating pressure 31 MPa, use of either syngas or pure hydrogen as reagent gas, recirculation of the gas after purification, recirculation of selected cut to be used as oil carrier.

The flow diagram of the PDU is presented in Figure 1. A modular concept was used in designing the system. Mass balances can be closed over each module and also changes can be made within each module without disturbing the others. The system can be regarded as a combination of 6 modules:

- slurry preparation and feed
- fresh gas manifolds
- reactor
- led-down and product collection
- gas scrubbing
- gas recirculation.

The slurry preparation module is comprised of a slurry preparation tank, provided with a multi-blade stirrer recirculation loop, electric heater and connection to inert gas and to the slurry feed tank. The loading of this tank is made batch-wise, each batch being enough for 7 hours unit operation at a feeding rate of 6 kg of slurry/hour. From this vessel, the slurry is transferred to the slurry feed tank. The construction of this tank is similar to that of the preparation tank, but for the purpose of feed rate monitoring, this vessel is suspended on a load cell. The slurry is continuously recirculated at a sufficient rate to keep the coal in suspension by a low-pressure recirculation pump. The slurry is heated under inert blanket at all times. A high-pressure piston pump takes suction from the recirculation line at a pressure of 50-70 KPa. The slurry feed rate can be changed by adjusting the stroke.

The fresh gas feed module includes a separate manifold for each of the following gases: hydrogen, carbon monoxide, nitrogen and helium; each provided with pressure regulators and flow meters. The fresh gas is supplied from commercially available cylinders. The flow of each individual gas is regulated by micrometric valves.

The reactor module comprises the slurry preheater, the gas preheater, the upflow reactor, hot high-pressure receiver, the condenser and the backpressure regulator.

All the heaters are electric and temperatures are automatically controlled. The pressures are regulated by spring-loaded pressure regulators.

The slurry preheater is a coil imbedded in a heat transfer cement block, heated by an electric heater. The gas preheater is a larger diameter pipe, provided with strip heaters.

The possibility of splitting the gas flow before and/or after the preheater is a built-in option serving a double purpose: To prevent slurry preheater coking in case of a temperature run in the preheater and to help cool the in-flowing slurry in either the preheater or the reactor in case of a temperature run in the reactor.

The reactor is an empty, upflow tubular reactor, built with modified high-pressure Bridgeman closures, provided with 4 heating zones, individually controlled by thermocouples placed directly in the flow.

The reaction mixture leaving the reactor at the reaction temperature and pressure reaches the high temperature, high-pressure receiver where the first flash takes place. The vessel is tubular, provided with the same type of closures as the reactor. The liquid level is kept constant with the help of a γ-ray level indicator.

The vapor phase comprised of unreacted H_2, CO, H_2O and produced CO_2, H_2S, NH_3, and low boiling hydrocarbons leave the hot high-pressure receiver at the top, go through a water-cooled condenser where the light fraction and the water are condensed and reach the cold high-pressure receiver. The non-condensed gases are separated at the top and led to the scrubber and the low boiling product containing water, some sulphur and ammonia products and light hydrocarbons are collected periodically.

The liquid phase from the hot high-pressure receiver is evacuated via the pressure let-down module. The heavy product pressure let-down module is constituted of two separate trains, each comprising two sequentially operated on/off actuated valves with a small intermediate vessel in-between. The timing of the valves is manually adjusted, based on the indication of the γ-ray indicator, the feeding rate and reaction parameters.

The slurry stream can be directed on either of the let-down trains while the other one is completely isolated allowing eventual repair work to proceed.

The heavy liquid stream off the let-down system is collected in a hot receiver where the last of the dissolved gas, water vapors and products boiling lower than 200°C are leaving the liquid phase in the process of the pressure release, and after being cooled in a condenser the condensate is periodically collected and analysed as medium product.

The liquid phase evacuated hourly from the hot low-pressure receiver contains the ash and unreacted coal along with heavy product and recuperated solvent fraction.

The gas scrubber module was designed for operation at ambient temperature with 4 MPa. The scrubber itself is a packed column where the gas moves in counter current with caustic solution. The absorbent solution is circulated with a pump to the top of the scrubber and is periodically analysed. The solution leaving the scrubber is kept in a closed tank connected to the exhaust system in order to allow the dissolved gases to be monitored and vented.

The gas recycle module involves a scrubbed gas storage tank kept at 4 MPa by a system of spring-loaded pressure regulators. A diaphragm compressor takes suction from this tank and discharges the gas mixture at high-pressure into a buffer tank which serves as gas feed tank. Recycle gas composition is analysed hourly and calculated ratios of fresh make-up gases are added into the compressor's suction to achieve the desired H_2/CO ratio. The recycle module was not used in this program.

In its present configuration, the PDU can operate in a variety of modes due to particular built-in features which were included in the design in view of confering the great versatility required from a testing facility.

The following conditions can be changed and their influence can be studied in relation to conversion of coal or liquid products yields:

- solids content in the initial slurry
- slurry feeding rate
- gas flow and gas/coal ratio
- initial gas consumption
- temperature of the feed
- temperature and pressure of the reaction module
- retention time at reaction temperature
- type and composition of the solvent
- temperature of the hot high-pressure receiver
- temperature of the hot low-pressure receiver
- liquid flow through the scrubber
- retention time in the hot high-pressure receiver
- temperature and velocity of the slurry through the preheater
- once-through catalysts can be added to the coal or solvent.

Testing at one given set of conditions implies operating the system at those conditions for enough time to satisfy the steady state condition after which data are collected for a period of 4 to 8 hours. The recorded data are: pressures and temperatures in all critical points, liquid and gas flows in and out of the unit, gas composition in the gas feed storage tank, before and after the scrubber and in the gas exhaust lines.

1.1 PROCEDURES

1.1.1 FEEDSTOCK PREPARATION AND CHARACTERIZATION

Approximately 500 Kg of each coal (Monkmand and Greenhills) were obtained by the client from B.C. mines. The Greenhills coal came from Sparwood mine (50% from seam No. 1, 25% from seam No. 7 and 25% from seam No. 10) The coal samples were shipped first to B.C. Research (BCR). Samples were taken for tubing bomb tests at BCR and the remainder shipped to Ontario Research Foundation (ORF) for stage grinding to 100 mesh. The ground coal was sealed in plastic bags and delivered to SBRG for continuous testing.

Proximate and ultimate analyses were done by SBRG on the two coals.

The moisture was determined both by xylol distillation and according to ASTM 3172-73. The ASTM method for the determination of volatile content was slightly modified to avoid 'popping' and to suit the equipment available. The modification related to the rate of heating to 650°C which was slower (samples reached 650°C in 35 minutes instead of 6 minutes as per ASTM 3174-73). After reaching 650°C the samples were moved into a second furnace at 950°C for 8 minutes.

The elemental composition was conducted on moisture free samples using a Perkin-Elmer Model 240 Elemental Analyzer. Sulfur was determined with a Leco model SC32 sulfur determinator. The ash was analysed for the following metals: Al, Ca, Co, Cu, Fe, K, Mg, Mo, Na, Ni, Sn, V.

Anthracene oil from Crowley Tar was catalytically hydrogenated in the SBRG continuous unit and used as solvent after filtering out the catalyst. The solvent was analysed for C, H, N, and S content, water and ash content and boiling range. The ash was further analysed for Co and Mo content to determine the amount of catalyst remaining in the solvent after filtration.

Commercial Co-Mo /γ-Al_2O_3 catalysts were obtained from American Cyanamid and stage ground to - 100 mesh by ORF. Solvent hydrogenation was done with HDS-2 and experiment number 13 (Monkman coal) was done with HDS 1442. The catalysts are quite similar and different catalysts were used for reasons of availability only. The analysis obtained from American Cyanamid is presented in section 1.3.1.

Red mud from the Alcan Arvida plant was used as catalyst in experiment number 14 (Green Hills coal). About 240 kg of red mud slurry were received and treated by settling and air drying under infrared lamps to remove the water. The dry mud was very fine (-200 mesh) so that no grinding or further preparation was necessary. The dried mud was analysed for water content only. A mineral analysis was obtained from Kilborn.

The sulfur added to the feedstocks in experiments number 13 and 14 was obtained from BDH.

Very dry grade H_2 from Specialty Gas was used.

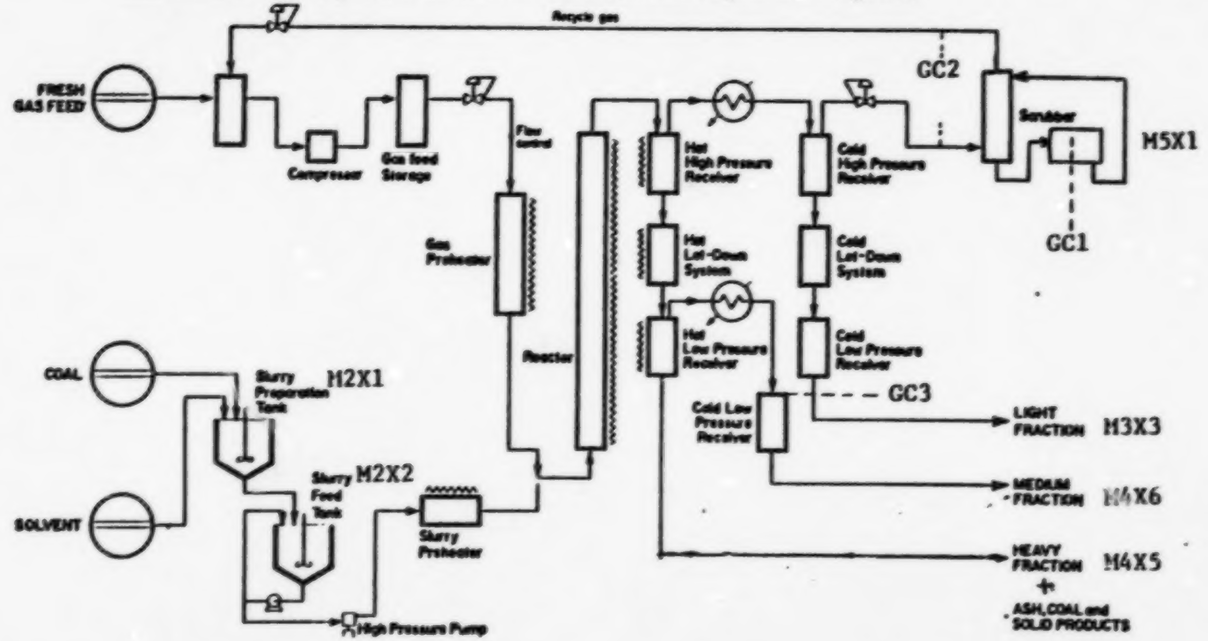
1.1.2 CONTINUOUS TESTING

The coal, solvent, catalyst and sulfur were mixed in the feed preparation vessel to make up 45 kg batches of slurry. This slurry was heated to 70°C and continuously mixed and recirculated to maintain a homogeneous mixture. When required, the slurry was transferred by gravity to the feed vessel where it was maintained at 70°C and continuously mixed and recirculated. Feed samples were taken from the feed vessel (M2X2) at intervals of two hours and analysed for water content by xylol distillation and for solids content by THF filtration.

Pure hydrogen was used directly in a once through mode - no recycling of the gases was used.

FIGURE 1:

Schematic flow diagram of the continuous flow coal liquefaction system


Recycle gas
FRESH GAS FEED
Compressor
Gas feed Storage
Flow control
Gas Preheater
Reactor
COAL
SOLVENT
Slurry Preparation Tank
M2X1
Slurry Feed Tank
M2X2
Slurry Preheater
High Pressure Pump
Hot High Pressure Receiver
Hot Let-Down System
Hot Low Pressure Receiver
Cold High Pressure Receiver
Cold Let-Down System
Cold Low Pressure Receiver
Cold Low Pressure Receiver
Scrubber
M5X1
GC1
GC2
GC3
LIGHT FRACTION
M3X3
MEDIUM FRACTION
M4X6
HEAVY FRACTION
M4X5
+
ASH, COAL and SOLID PRODUCTS

Product recovery
    Liquids
        Light fraction - M3X3
        Middle fraction - M4X6
        Heavy fraction - M4X5

Gases
Scrubber off gas - M5X1 - GC1
Scrubbed gas - GC2
Product off gas - M4X5 - GC3

The product streams were as follows:

heavy product - M4x5 vessel

This includes all condensed products not flashed off the hot high-pressure receiver or distilled or entrained from M4X5 to M4X6. This product includes all ash and unconverted solids, most of the oils and some water. Usually it makes up 90 wt% of the total product.

middle product - M4X6 vessel

All condensibles distilled from M4X5 are recuperated in M4X6. This stream will contain some water and light organics and makes up 2 to 5% of the total product.

light product - M3X3 vessel

The condensibles flashed from the hot high-pressure receiver are recuperated in this vessel. The stream contains water and light organics and usually makes up about 1% of the total products.

non-condensible gases

The product gas is sampled before and after scrubbing. Also the off gas from the M4X5 vessel and scrubber recirculation tank (M5X1) are sampled periodically. Volume measurements are taken of the scrubbed gas and both off gases.

The heavy product was sampled every hour and analysed for THF insoluble content by filtration. The reaction was assumed to be at steady state when all operating conditions were constant and THF solids content was constant to within 1%. Once steady state was reached the products were placed in new receivers and kept separate from any interim products. M4X5 and M4X6 were emptied every 45 minutes and the product weighed. M3X3 was emptied every two hours and the product weighed.

1.1.3 PRODUCT ANALYSIS

The total heavy product (M4X5) obtained in an experiment was well mixed and representative samples taken for the following analyses:

- water content by xylol distillation
- THF insoluble content by filtration
- distillation into 4 fractions
  - <200°C
  - 200 - 325°C
  - 325 - 495°C
  - 485°C+
- C H N S on each fraction
- O by n-activation on some of the fractions
- ash, Ni, Mo, Fe, Co on THF insolubles

Some of the heavy product was entrained into M4X6 during the experiment number 13 and so this was analysed as described above. During experiment number 14 M4X6 was not operational and so the middle product was contained in the heavy product.

The liquid product from M3X3 was separated into a water fraction and an organic fraction by liquid extraction in a separatory funnel and the organic fraction was analysed for C, H, N and S content.

The gas samples were analysed using a Carle gas chromatograph model AGC 111-SX (application no. 158B). The following components were separated: H_2, CO, CO_2, CH_4, C_2H_6, C_3H_8 and H_2S. The gas was not filtered or cleaned before analysis to allow better accuracy in the H_2S determination.

1.2 TESTING CONDITIONS

1.2.1 SOLVENT HYDROGENATION

The raw anthracene oil was catalytically hydrogenated in the continuous unit in order to increase its hydrogen donor capabilities. The anthracene oil and ground catalyst were mixed in the feed preparation vessel and fed to the reactor under the conditions given below. The 'heavy product' from the M4X5 vessel was then filtered through 200 mesh screen and glass wool to remove the catalyst and the filtrate was used as solvent for experiments number 13 and 14. The products in M4X6 and M3X3 were not mixed with the solvent since they were too light. At the end of the experiment, all of the steady state product from M4X5 (about 190 Kg was mixed together to give a uniform solvent for further experimentation. This product was analysed for water content, elemental composition, distillation, ash content and Co & Mo content in the ash.

The catalyst tended to settle in the slurry lines since the oil was not viscous enough to keep it in suspension. This caused some feeding problems but the flow was maintained for 36.2 hours. The product gas contained a large NH_3 fraction for which the gas chromatograph was not calibrated. This made product gas analysis impossible but the main purpose of the experiment, that is, generating a hydrogenated solvent, was satisfied.

1.2.2 COAL HYDROGENATION

The Monkman coal was slurried with the hydrogenated solvent, Co-Mo catalyst and sulphur. This slurry was fed into the reactor at a rate of 5.8 Kg/hr for a total of 11.6 hours. Steady state was achieved after 5.3 hours and was maintained for the last 7.3 hours (experiment number 13).

About 50 Kg of anthracene oil were then passed through the reactor to flush the system during 6 hours before the next experiment.

The Greenhills coal was then slurried with hydrogenated solvent, red mud as catalyst, and sulphur. The initial attempt at experiment number 14 was unsuccessful because of a failure in the Badger-Meter let down valve and problems with the feed module which caused erratic slurry flow. These problems were solved, and experiment number 14 was carried out smoothly. The slurry was fed in at a rate of 6.3 Kg/hr for a total of 10.1 hours and steady state was obtained after 5 hours and was maintained for the last 5.0 hours.

A new batch of anthracene oil was added to the feed vessel for flushing of the system and the unit was brought down.

Operating conditions for experiments number 13 and 14 are given in Table 3.

1.3 RESULTS AND DISCUSSION

1.3.1 FEEDSTOCK CHARACTERISTICS

Proximate, ultimate and metals analysis done by SBRG on the two coals are presented in Table 4. Samples taken from the top and bottom of the coal drums used showed no significant difference in the analyses.

Manufacturers specifications for the catalysts used during the run and presented in Table 5.

1.3.2 SOLVENT HYDROGENATION

The hydrogenated solvent, that is, the filtered M4X5 product from the catalytic hydrogenation of anthracene oil, made up about 75.5 wt % of the anthracene oil fed to the unit. The remaining 24.5 wt % was made up of gas and lighter products found in M4X6 and M3X3 and was not added to the solvent. A comparison of the original anthracene oil and the hydrogenated solvent is presented in Table 6.

The H content has been increased from 5.7 wt % to 7.5 wt %. The sulphur content in the solvent is negligible, almost all of the sulphur was reacted to H_2S. An analysis of the ash from the solvent shows that some of the catalyst remained in the solvent after filtration. Analysis found cobalt concentrations of 0.07 ppm and molybdenum concentractions of 0.24 ppm in the solvent. At these concentrations the metals are not in sufficient quantities to significantly catalyze the coal hydrogenation experiments.

1.3.3 COAL LIQUEFACTION TESTS

The material balances for the coal hydrogenation experiments are given in Table 7. The output data is prorated to give 100% mass balance. Actual mass balances were within 3%. The ash OUT was assumed to be equal to ash IN, a direct determination of the output ash being difficult because of non-homogeneity of the residual product and the small quantities (1 gram) used in ash determination.

The conversions, total liquid yields and distillable product yields obtained in the two tests, are presented in Table 8.

The coal conversion is calculated on the basis that THF insoluble material found in the product is comprised of unreacted coal plus ash plus catalyst.

The calculation of the liquid yields (total liquid and distillable yield) must take in to account the presence in the final product, of the solvent introduced. The tests being done in the once through mode, the solvent constitutes up to 70% of the final liquid product. In order to calculate the coal derived liquid yields an assumption on the fate of the solvent must be made, and this assumption has a great bearing on the results obtained. The Table 8 presents the conversion and yields obtained at the conditions tested calculated for two limiting assumptions:

Assumption A

The solvent does not change in presence of coal and catalyst and at the conditions tested, in other words that SOLVENT OUT SOLVENT IN or LIQUID PRODUCT - SOLVENT IN COAL DERIVED LIQUID.

This assumption represents a limiting case resulting in low distillable yields (19% and 17.7%), lower than the real yields probably are.

Assumption B

All hydrocarbon gases obtained in the test (C_1 to C_3) result from cracking of the solvent. In this case, SOLVENT OUT SOLVENT IN - (C_1 & C_2 & C_3) and COAL DERIVED LIQUID TOTAL LIQUID - SOLVENT - (C_1 & C_2 & C_3). This assumption almost doubles the distillable yields as shown in Section B and show how sensitive the calculations are to solvent behaviour.

In fact, as a result of Assumptions made in B, liquid yields calculated appear to be higher than the conversions and the gas yields are negative. Although this is not possible in reality, it does demonstrate the consequences of applying this limiting assumption on the calculations of distillable yields.

The yields from an eventual process using recycle solvent would probably be somewhere between these two calculations. The conversions and yields obtained by us are in agreement with the results reported by researchers at EXXON (see Appendix A)., where, for coals having a volatile matter content similar to the coals tested (around 20%), the distillable yield obtained was not higher than 20%.

The low distillable yields in relation to the much higher total liquid yields indicates that:

- "solubilization" of some macromolecular structures present in coal is achieved up to levels of total liquid yield comparable to other coals that SBRG has tested under similar conditions.
- hydrocracking of the solubilized product into similar (distillable) fractions is not achieved at any large extent under the conditions used.

The results thus obtained should be interpreted in the following way:

- under the conditions selected for these tests, the two coals appeared not to be very reactive, since their distillate yields are low.
- the limiting step seems to be the extent of hydrocracking of the high molecular weight compounds (initially solubilized) to lower molecular weights.

Our main conclusion is that a different approach is needed for this type of cretaceous coals. The following options should be considered:

1) Prior to treatment of these coals with suitable agents can lead to physico-chemical and structure changes which may enhance their reactivity.
2) Using harsher experimental conditions (higher T, P and G:L ratios gas to liquid) might facilitate the break-down of some quasi-inertinitic material and at the same time, induce further hydrocracking of the readily solubilized high molecular weight materials.
3) Severe catalytic hydrotreating of the high molecular weight liquids (THF solubles). This implies a two-step liquefaction process.

Analyses of the liquid products are presented in Tables 9 and 10. The first part of the tables shows the product stream distribution and the breakdown into solids, water and organic liquid fractions. In experiment 13 there was a heavy liquid carry over from M4X5 to M4X6 so that the latter vessel did not contain only the middle product for which it was designed but a heavy product instead. This M4X6 heavy product was analyzed separately and fully characterized. In experiment 14 a plug in the line prevented any product from reaching M4X6 and all liquid products were found in M4X5 and M3X3.

Distillation cuts and elemental composition are given in the lower part of Tables 9 and 10. Hydrogen content decreases significantly for the heavier cuts whereas the heteroatoms tend to concentrate in the distillation residue. Sulphur for example is found almost exclusively in the distillation residue and as H_2S. The behaviour of the two coals shows no marked difference. The M3X3 product was not distilled since the quantities were small. Typically this product boils in the range 55°C - 290°C.

Tables 11 and 12 give the output gas analysis for experiments 13 and 14. The volume of gases dissolved in the heavy product (M4X5) and scrubber water (M5X1) are seen to be quite significant. Gases dissolved in the light product (M3X3) were not determined separately they appear on the table as part of the M4X5 off gases. There was almost no CO or CO_2 production which is due to the low oxygen content of the coals and solvent. The H_2S is concentrated almost completely in the M4X5 off gases. The $C_1 - C_3$ yield is seen to be quite important in terms of hydrogen consumed.

Tables 13 and 14 give the elemental mass balances for the experiments. The carbon and hydrogen balances obtained are within 5%.

TABLE 1: DESIGN SPECIFICATIONS FOR PDU

Slurry feed rate	10 kg/hr
Gas feed rate	1200 m^3/t coal
Slurry preheat temp.	300°C
Gas preheat temp.	400°C
Reactor temp.	475°C
Reactor press.	31 MPa
Scrubber press.	4 MPa
Retention time at reaction temp.	60 min

TABLE 2: REACTION CONDITIONS FOR ANTHRACENE OIL HYDROGENATION

Oil	: anthracene oil from Crowley Tar
Catalyst	: HDS-2 Co-Mo from American Cyanamid stage ground to - 100 mesh
Reactor pressure (MPa)	: 20.7 (3000 psi)
Temperatures (°C)	
reactor	: 440
slurry preheater	: 290
gas preheater	: 335
hot receiver	: 180
cold receiver	: 30
Flows	
slurry (kg/h)	: 7.1
gas (L/h)	: 1400
Catalyst concentration	: 2.5% catalyst

TABLE 3: OPERATING CONDITIONS FOR COAL LIQUEFACTION TESTS

Experiment No.			13	14
Coal			M.K.M.	G.H.
Catalyst			HDS1442	R.M.
Reactor Pressure		MPa	20.7	26.6
Temperature	Reactor	°C	440	440
	Slurry Preheater	°C	285	290
	Gas Preheater	°C	350	345
	Hot receiver	°C	200	175
	Cold receiver	°C	15	19
Flows	Slurry	Kg/hr	5.80	6.30
	mf coal	Kg/hr	1.90	2.11
	gas	L/hr	2450	2060
Coal Composition	moisture	% wt	1.05	1.15
	ash	% wt	17.75	14.10
	maf coal	% wt	81.21	84.91
Reactants Ratios	coal as recd/solvent	Kg/Kg	1.0/2	1.06/2
	maf coal/solvent	Kg/Kg	.81/2	.90/2
	H2/coal maf	L/Kg	1043	726.6
Catalyst	Hourly feed	Kg/hr	0.033	0.082
	Cat/coal maf ratios	% wt	2.0	5.0
Sulfur	Hourly feed	Kg/hr	0.005	0.041
	S_2/coal maf	% wt	0.3	2.21

TABLE 4: ANALYSIS OF GREENHILLS AND MONKMAN COALS DONE AT SBRG

Analysis	Green Hills	Monkman
Proximate (% wt)		
Moisture		
oven	1.2	1.1
xylol distillation	(1.4)	(1.6)
Ash	14.1	17.7
Volatile	19.6	15.9
Fixed carbon	65.1	65.3
Ultimate (% wt)		
C	75.0	73.0
H	4.64	4.43
N	1.53	1.28
S	0.39	0.38
O by difference	4.34	3.21
Metals (ppm)		
Al	13 500	6 600
Ca	1 410	1 750
Co	3.0	3.4
Cu	40.0	6.9
Fe	6 470	3 300
K	810	1 070
Mg	440	840
Mo	2.0	2.0
Na	31	248
Ni	7.7	9.9
Sn	<.05	<.05
V	30	34

TABLE 5: * CATALYST SPECIFICATIONS

AERO HDS-1442B (American Cyanamid)

used in solvent hydrogenation
manufacturers specifications

MoO_3(wt %)	14.5 - 16.5
CoO	3.2 - 4.0
SiO_2	1.0 max.
SO_4	0.8 max.
Na_2O	0.04 max.

AERO HDS-2 (American Cyanamid)

used in experiment number 13 (Monkman)
manufacturers specifications

MoO_3(wt %)	15.4
CoO	3.2
SiO_2	0.1
SO_4	0.3
Na_2O	0.03

Arvida Red Mud (Alcan)

used in experiment number 14 (Green Hills)
manufacturers specification

Fe_2O_3(wt %)	33
TiO_2	6
SiO_2	17
Al_2O_3	20
CaO	1
Na_2O	9
H_2O (wt %) (after drying at SBRG)	0.9

TABLE 6: COMPARISON OF HYDROGENATED SOLVENT AND ORIGINAL ANTHRACENE OIL

	Anthracene Oil	Hydrogenated Solvent
	wt %	
Water	nil	nil
Distillation cuts		
< 200°C	3.4	3.98
200 - 325°C	40.0	61.47
325 - 475°C	51.0 (to 412°C)	33.43
475°C	5.6 (412°C)	1.12
Elemental		
C	91.1	91.1
H	5.70	7.52
N	0.97	0.43
S	0.62	.005
O by difference	1.61	0.9
Ash	nil	.05
Ash Analysis		
Co (ppm)	not done	144
Mo (ppm)	not done	480

TABLE 7: MATERIAL BALANCES FOR CONTINUOUS TESTING OF MONKMAN AND GREEN HILLS COALS

INPUT (kg)

	EXPERIMENT 13	EXPERIMENT 14
Water	0.15	0.12
Ash	2.48	1.49
maf coal	11.37	9.06
Solvent	27.99	20.22
Hydrogen	1.29	1.17
Catalyst	0.24	0.409
Sulphur	0.034	0.206
Total kg	43.56	32.68

OUTPUT (kg)

	13	14
Water	0.45	0.23
Ash*	2.73	1.90
THF filtration insoluble - ash	1.98	2.2
Organic Liquid (THF solubles - water)	35.74	26.16
Hydrogen	0.921	0.843
H_2S	0.006	0.132
Other gases	1.734	1.214
Total	43.56	32.68

* includes the catalyst and sulfur unreacted to H_2S

TABLE 8: CONVERSION AND YIELDS FROM ONCE THROUGH CONTINUOUS TESTING USING HYDROGENATED ANTHRACENE OIL AS SOLVENT

A. Assumption: solvent out = solvent in

% maf coal:	MONKMAN	GREENHILLS
Conversion $^{1)}$	82.8	76.6
Liquid Yield $^{2)}$	68.2	65.6
Gas Yield $^{3)}$	12.0	9.8
Water Yield $^{4)}$	2.6	1.2
Distillate Yield $^{5)}$	19.00	17.7
H_2 Consumption $^{6)}$	3.24	3.59

B. Assumption: solvent out = solvent in − (C1 + C2 + C3)
(All C1, C2, C3 gas result from solvent cracking)

% maf coal	MONKMAN	GREENHILLS
Conversion $^{1)}$	82.8	76.6
Liquid Yield $^{2)}$	83.3	79.3
Gas Yield $^{3)}$	(3.1)	(3.9)
Water Yield $^{4)}$	2.6	1.2
Distillable Yield $^{5)}$	34.04	33.84
H_2 Consumption $^{6)}$	3.24	3.59

1) conversion = (m.f. coal + catalyst − THF insolubles + S not in gas) ÷ maf co
2) liquid yield = (THF solubles − water out − solvent out) ÷ maf coal
3) gas yield = (gas out − gas in − S in H_2S) ÷ maf coal
4) water yield = (water out − water in) ÷ maf coal
5) distillate yield = (distillables − solvent out) ÷ maf coal
6) H_2 consumption = (H_2 in H_2 out) maf coal

TABLE 9: ANALYSIS OF LIQUID PRODUCTS FROM EXPERIMENT NO. 13 (MONKMAN COAL)

Product Stream	THF Insolubles % wt	THF Insolubles kg	Water (xylene method) % wt	Water (xylene method) kg	Organic Liquids % wt	Organic Liquids kg	Total kg
M4x5	12.1	4.35	0.12	0.04	87.8	31.56	35.95
M4X6	9.2	0.36	0.5	0.02	90.2	3.49	3.87
M3X3	0		36.1	0.39	63.9	0.69	1.08
Total(kg)		4.71		0.45		35.74	40.90

Distillation cuts and elemental composition

distillation cut	ash + catalyst	C	H	N	S	O by diff	total
	% wt						
M4X5 product IBP: 120°C							
H_2O			11.1			88.9	0.12
<200°C		88.9	10.6	0.67	.05	-	2.78
200 - 325°C		89.9	8.76	0.67	.05	0.67	47.2
325 - 475°C		92.5	6.93	0.50	.05	0.07	23.1
475°C+	25.75	66.71	3.91	1.35	0.48	1.80	26.8
							100.00
M4X6 product IBP: 100°C							
H_2O			11.1			88.9	0.5
<200°C		88.9	11.0	0.39	.05	-	3.19
200 - 325°C		90.9	8.47	0.33	.05	0.3	50.47
325 - 475°C		91.6	7.00	0.53	.05	0.87	20.24
475°C+	27.20	64.2	3.31	2.30	0.40	1.60	25.60
							100.00
M3X3 product							
H_2O			11.1			88.9	36.06
organic		88.9	10.7	0.40	.05	0	63.94
							100.00

TABLE 10: ANALYSIS OF LIQUID PRODUCTS FROM EXPERIMENT NO. 14 (GREENHILLS COAL)

Product Stream	THF Insolubles		Water (xylene method)		Organic Liquids		Total
	% wt	kg	% wt	kg	% wt	kg	kg
M4x5	13.85	4.21	0	0	86.15	26.2	30.41
M3X3			28.2	0.24	71.8	0.61	0.85
Total(kg)		4.21		0.24		26.81	31.26

Distillation cuts and elemental composition

M4x5 product
IBP: 110°C

distillation cut	ash + catalyst	C	H	N	S	O by diff	total
	% wt						
H_2O							0
<200°C		89.1	10.1	0.55	.05	-	3.25
200 - 325°C		90.8	8.18	0.40	.05	0.62	45.26
325 - 480°C		92.2	6.61	0.66	.05	0.53	22.28
480°C+	21.94	68.77	4.16	2.03	1.05	2.05	29.21
							100.00

M3X3 product

		C	H	N	S	O	total
H_2O		n.d.	11.1	n.d.	n.d.	88.9	31.09
organic		88.5	10.0	0.52	0	0.98	69.91
							100.00

TABLE 11: OUTPUT GAS ANALYSIS FOR EXPERIMENT NO. 13

OUTPUT GAS	SCRUBBED	PERCENT VOLUME M4X5	M5X1	TOTAL
H2	.898	.871	.792	.892
CO	0.000	0.000	0.000	0.000
CO2	.001	.001	0.000	.001
CH4	.036	.053	.113	.040
C2H6	.014	.030	.047	.016
C3H8	.051	.039	.048	.050
H2S	0.000	.006	0.000	.000
VOLUME(L1)	10418.0	694.0	452.0	11564.0

	SCRUBBED	PERCENT WEIGHT M4X5	M5X1	TOTAL
H2	.355	.319	.229	.346
CO	0.000	0.000	0.000	0.000
CO2	.005	.005	0.000	.005
CH4	.115	.154	.262	.125
C2H6	.084	.167	.203	.096
C3H8	.441	.317	.306	.426
H2S	0.000	.038	0.000	.002

WEIGHTS TABLE (KG)

H2	.835	.054	.032	.921
CO	0.000	0.000	0.000	0.000
CO2	.012	.001	0.000	.013
CH4	.271	.026	.037	.333
C2H6	.198	.028	.028	.255
C3H8	1.037	.054	.043	1.133
H2S	0.000	.006	0.000	.006
GRAM MOLES	464.8	31.0	20.2	515.9
WEIGHT(KG)	2.353	.169	.139	2.661

ELEMENTAL COMPOSITION (KG)

HYDROGEN	1.131	.076	.054	1.261
CARBON	1.213	.086	.085	1.384
OXYGEN	.009	.001	0.000	.010
SULPHUR	0.000	.006	0.000	.006

TABLE 12: OUTPUT GAS ANALYSIS FOR EXPERIMENT NO. 14

OUTPUT GAS	PERCENT VOLUME SCRUBBED	M4X5	M5X1	TOTAL
H2	.932	.381	.659	.898
CO	0.000	0.000	.002	.000
CO2	0.000	.002	.005	.000
CH4	.036	.195	.132	.047
C2H6	.013	.086	.051	.018
C3H8	.018	.174	.136	.029
H2S	0.000	.161	.015	.008
VOLUME(L)	9963.0	524.0	314.0	10801.0

	PERCENT WEIGHT SCRUBBED	M4X5	M5X1	TOTAL
H2	.511	.039	.113	.385
CO	0.000	0.000	.005	.000
CO2	0.000	.005	.018	.002
CH4	.159	.159	.180	.161
C2H6	.109	.131	.131	.115
C3H8	.220	.388	.509	.275
H2S	0.000	.278	.044	.060

WEIGHTS TABLE (KG)

H2	.829	.018	.018	.865
CO	0.000	0.000	.001	.001
CO2	0.000	.002	.003	.005
CH4	.258	.073	.030	.361
C2H6	.177	.060	.021	.259
C3H8	.356	.179	.084	.618
H2S	0.000	.128	.007	.135
GRAM MOLES	444.5	23.4	14.0	481.8
WEIGHT(KG)	1.620	.461	.164	2.245

ELEMENTAL COMPOSITION (KG)

HYDROGEN	.993	.088	.046	1.127
CARBON	.627	.250	.109	.986
OXYGEN	0.000	.002	.003	.004
SULPHUR	0.000	.121	.007	.127

TABLE 13: ELEMENTAL MASS BALANCE FOR EXPERIMENT NO. 13 (MONKMAN COAL)

FEED	ash + catalyst	C	H	N	S	O	total
				KG			
H_2O			.01			.14	0.15
m.f. coal	2.72	10.10	0.61	0.18	0.087	0.45	14.15
Solvent		25.50	2.10	0.12	0.01	0.25	27.98
H_2			1.29				1.29
Total	2.72	35.60	4.01	0.30	0.097	0.84	43.57
PRODUCT							
Heavy Product							
<200°C		1.03	0.12	0.011	.044	0	1.16
200 - 325°C		17.03	1.65	0.116	.006	0.13	18.94
325 - 475°C		8.40	0.63	0.044	.004	0.06	9.14
475°C+	2.73	7.05	0.41	0.15	0.054	0.19	10.57
Light Product		0.61	0.07	0.003	.001		0.68
H_2O			0.05			0.40	0.45
Gases							
M4X5		0.09	0.08	n.a.	0.006	0.001	0.18
M5X1		0.09	0.05	n.a.	0	0	0.14
Scrubbed		1.21	1.13	n.a.	0	0.009	2.35
Total	2.73	35.51	4.19	0.23	0.073	0.79	43.61

TABLE 14: ELEMENTAL MASS BALANCE FOR EXPERIMENT NO. 14 (GREEN HILLS COAL)

FEED	ash + catalyst	C	H	N	S	O	total
	KG						
H_2O			0.01			0.11	0.12
m.f. coal	1.90	7.91	0.49	0.16	0.247	0.46	11.17
Solvent		18.42	1.52	0.09	0.01	0.18	20.22
H_2			1.17				1.17
Total		26.33	3.19	0.25	0.257	0.75	32.68
PRODUCT							
Heavy Product							
<200°C		0.86	0.10	0.005	0.001	0	0.96
200 - 325°C		12.19	1.10	0.054	.007	.08	13.43
325 - 475°C		6.09	0.44	0.044	.003	.035	6.61
475°C	1.90	5.96	0.36	0.176	0.091	0.178	8.66
Light Product		0.53	0.06	0.003	0	0.007	0.60
H_2O			0.014			0.216	0.23
Gases							
M4X5		0.25	0.09	0	0.121	0.002	0.46
M5X1		0.11	0.05	0	0.007	0.003	0.17
Scrubbed		0.63	0.99	0	0	0	1.62
Total	1.90	26.62	3.20	0.287	0.230	0.521	32.76

Appendix A:

EDS COAL LIQUEFACTION PROCESS DEVELOPMENT ANNUAL TECHNICAL PROGRESS REPORT JULY 1979 - JUNE 1980

4.6 "At Large" Coal Screening

Tubing bomb liquefaction experiments were performed on eleven "at large" coals to select coals for possible further study, to enlarge the bench liquefaction data base and for use in improving the coal screening correlation (July, 1978-June, 1979, EDS Annual Technical Progress Report [FE-2893-35]). Of these eleven samples, four were from the Rocky Mountain region, three were from the Western Interior region and the remaining four were from the Appalachian region.

Liquefaction experiments were conducted in triplicate at a single condition (840°F/40 minutes/1500 psig). The solvent(tetralin)-to-coal ratio was 1.6/1. Liquefaction conversion was calculated from cyclohexane insolubles and by ash balance. The gas make was obtained by water displacement. The water yield estimation was obtained by neutron activation analysis and oxygen balance.

Stepwise linear regression analysis was performed on these eleven coals. The independent variables included components of proximate and ultimate analyses and heteroatom-to-carbon ratios. Percent total sulfur and percent inorganic sulfur, all were expressed as wt % on a dry, mineral matter-free (DMF) coal basis; whereas mineral matter, percent total sulfur and percent inorganic sulfur were expressed as wt % on dry coal. The data base used is shown in Table 4-9.

TABLE 4-9

"AT-LARGE" COAL SCREENING DATA BASE

	ROCKY MOUNTAIN			WESTERN INTERIOR			APPALACHIAN				
State	NM	CO	UT	WY	MO	KS	IO	PA	PA	PA	PA
ASTM Rank	SBA-HVC	HVC	HVB	SBB	HVC-HVB	HVC-HVB	HVC-HVB	LV	HVA	MV-HVA	HVA
Proximate Analysis (Wt %)											
Mineral Matter (Dry)	8.84	9.30	5.86	7.02	11.78	14.8	5.97	7.95	6.05	9.07	42.90
Volatile Matter (DMMF)	44.96	42.97	40.09	50.11	41.02	40.07	48.42	18.95	36.49	23.91	37.38
Fixed Carbon (DMMF)	55.04	57.03	59.91	49.89	58.98	59.93	51.58	81.95	63.51	76.09	62.62
Ultimate Analysis (Wt % DMMF)											
Carbon	78.91	79.13	82.69	73.36	82.07	84.14	79.85	90.85	84.86	89.61	84.75
Hydrogen	5.67	5.51	5.68	5.14	5.48	5.71	5.70	4.80	5.61	4.96	5.60
Oxygen (Diff.)	13.54	13.11	9.57	19.68	9.27	6.76	9.55	2.59	7.10	2.67	7.32
Nitrogen	1.45	1.72	1.54	0.88	1.31	1.28	1.20	1.31	1.47	1.92	1.49
Sulfur (Organic)	0.42	0.52	0.50	0.93	1.85	1.82	3.67	0.37	0.87	0.82	0.57
Sulfur (Inorganic Dry)	0.11	0.10	0.20	0.06	1.94	1.64	2.17	0.42	0.67	0.30	1.45
Total Sulfur (Dry)	0.49	0.57	0.67	0.92	3.59	3.23	5.64	0.77	1.49	1.05	1.97
Tubing Bomb Liquefaction* (Wt % DAF)											
Conversion	53.0	47.8	45.3	59.5	53.7	49.1	60.2	11.2	41.5	18.0	42.9
Gas Make	8.7	8.3	6.1	13.1	6.3	5.7	6.7	3.1	5.3	3.7	4.9
CO_x	3.7	3.1	1.5	8.2	1.4	1.1	1.7	0.4	0.7	0.3	0.7
C_1-C_3	4.9	5.2	4.5	4.8	4.6	4.4	4.5	2.7	4.5	3.4	4.2
H_2S	0.1	0.0	0.1	0.1	0.3	0.0	0.5	0.0	0.1	0.0	0.0
Water	14.2	20.3	9.4	5.9	12.0	3.2	8.7	**	6.8	**	8.3
Hydrogen Consumption	3.5	3.6	2.9	3.2	2.5	2.6	2.8	3.4	2.4	**	2.6

---

*840°F/40 Min. 4 Wt % H_2

**Not Available.

The following observations are made based on the results of single variable correlation of the data:

- Not all components of the proximate and ultimate analyses are independent, e.g.,
  - Volatile matter (and hence fixed carbon) correlates reasonably well ($R > 0.80$) with oxygen-to-carbon ratio. Correlation of volatile matter is especially good ($R > 0.94$) with percent carbon and H/C ratio.
  - Percent carbon correlates well with percent oxygen and H/C and O/C ratios, as is expected from the results above.
- Tubing bomb conversion is most highly correlated with percent volatile matter (or fixed carbon), although, as expected from the above results, percent carbon and H/C ratio also correlate well with conversion.
- Total gas make correlates reasonably well ($R > 0.80$) with volatile matter (or fixed carbon) and especially well ($R \geq 0.95$) with percent carbon, percent oxygen and oxygen-to-carbon ratio.
- C_1-C_3 gas make is correlated highly with the H/C ratio and volatile matter (or fixed carbon).
- Mineral matter content is not strongly correlated with any other variable.

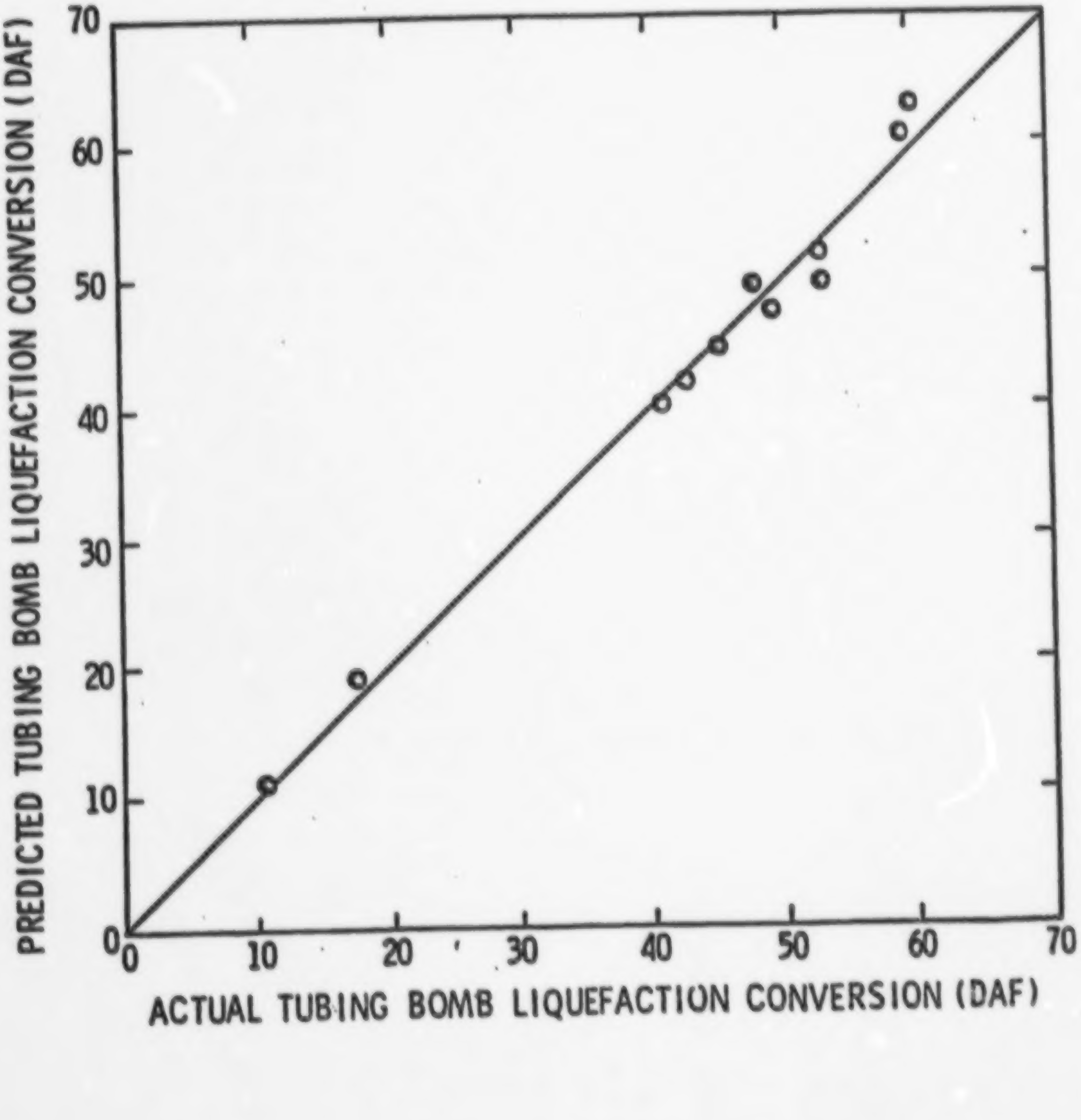
Although single variable correlation of conversion with volatile matter ($R=0.99$) and gas make with O/C ratio ($R=0.99$) were excellent for this sample of coals, stepwise linear regression of conversion, total gas make and C_1-C_3 gas yield with multiple variables was undertaken. Independent variables were chosen so that none used was highly intercorrelated with another.

The total sulfur content, in addition to volatile matter, was found to be significant when all non-correlated variables were regressed against conversion. No variable in addition to O/C ratio was found to reduce the variance in the gas yield correlation. Predicted conversion results are compared to the actual results for the eleven coals in Figure 4-1.

There is very little scatter about the parity lines in spite of the limited size and diversity of the coal data base used. The correlation obtained for conversion is similar to that obtained earlier with a more limited data base. The earlier correlation work was reported in the July, 1978-June, 1979, EDS Annual Technical Progress Report (FE-2893-35).

FIGURE 4-1

PARITY PLOT
TUBING BOMB CONVERSION



4.3 Correlation of Coal Conversion with Coal Properties

Stepwise multiple regression analysis of tube autoclave conversion data has been completed for six different coals and coal types. This is directed at developing a correlation expressing liquefaction conversion as a function of measurable coal properties and is similar to an effort underway at Pennsylvania State University.[1] Principal differences in the two studies include differences in feed coal preparation, conditions used for liquefaction, and in the definition of liquefaction conversion. As pointed out by Given, requirements for a valid statistical correlation include an extensive data base and a wide range of variability for each property studied. The work to date represents a modest beginning in this regard. More significance can be attached to the correlation as additional data are included.

Data used in the initial correlation work are given in Table 4-2. Independent variables include ultimate, proximate, and petrographic analyses. The dependent variable is taken to be the tubing bomb conversion (DAF) at 840°F and 40 minutes residence time, based on cyclohexane insoluble residue. Coals to date include two bituminous coals, two bituminous coal lithotypes, a subbituminous coal, and a lignite.

Results of single variable linear regression analyses are given in Table 4-3. Significance of each variable in explaining liquefaction behavior is expressed in terms of the percent of variation about the mean removed by regression. This is frequently referred to as the index of determination (R^2). Percent volatile matter was found to be the most significant variable, while percent mineral matter was the least significant. Only two variables were found to be of significance when all variables were simultaneously included in a stepwise linear regression analysis. These were percent volatile matter and percent total sulfur. The resulting index of determination was 96.8. Volatile matter was highly correlated with percent carbon, H/C atomic ratio, and percent exinite while total sulfur was not highly correlated with any other independent variables.

The resulting correlation was used to predict liquefaction conversion for several additional coals for which sufficient data for comparison was available. The result is shown in Figure 4-3. Predictions were made for four Gulf Coast lignites, two Pittsburgh Seam coals, two Australian coals (one a lignite and the other subbituminous), and an Illinois No. 6 coal. While some scatter does exist about the parity line, considering the limited number of coals and the great diversity of coal types included in the data base, the results are considered to be encouraging.

Several additional coals have been acquired for inclusion in the correlation data base in order to better cover the range of variables considered in the analysis. These are listed in Table 4-4. This will be a continuing activity throughout 1979 as more data become available.

---

[1] Given , P.H., et al., "Dependence of Coal Liquefaction Behavior on Coal Characteristics." 3. Statistical Correlations of Conversions in Coal Tetralin Interactions, FUEL, 57, February, 1978.

TABLE 4-2

COAL SCREENING DATA BASE

COAL	Pennsylvania Bituminous	Illinois #6 Bituminous	Wyoming Subbituminous	North Dakota Lignite	Illinois #6 Lithotypes	
					Vitrain	Fusain
Ultimate Analysis (Wt %, DMMF)						
Carbon	90.95	80.71	71.74	72.35	79.03	91.47
Hydrogen	4.84	5.15	5.44	4.66	5.44	2.99
Oxygen (Diff.)	2.22	9.31	21.50	21.54	11.64	3.88
Sulfur						
Total	0.87	5.64	0.53	0.86	3.82	2.16
Organic	0.55	3.39	0.32	0.56	2.48	1.12
Nitrogen	1.44	1.44	1.16	0.89	1.41	0.54
Proximate Analysis (Wt %)						
Mineral Matter (Dry)	7.8	14.5	8.8	6.2	6.6	14.07
Volatile Matter (DMMF)	19.3	43.9	48.4	46.5	44.0	17.5
H/C Atomic Ratio	0.65	0.78	0.90	0.78	0.83	0.41
Petrographic Analysis (Vol. %)						
Vitrinite	86.8	83.4	87.9	80.5	94.2	9.5
Exinite	0.0	5.4	4.3	2.7	2.3	0.4
Total Reactive Macerals	86.8	88.8	92.2	83.2	96.5	9.5
Tubing Bomb Liquefaction Conversion (DAF) (840°F/40 Min/4 wt % H_2)	9.7	56.6	51.9	47.5	59.3	19.3

TABLE 4-3

RESULTS OF SINGLE VARIABLE REGRESSION

Independent Variable	Percent of Variation About Mean Removed By Regression (R^2)
Volatile Matter	88.5
% Exinite	70.6
% Carbon	66.4
H/C Atomic Ratio	61.5
% Oxygen (Org.)	46.8
% Hydrogen	40.4
% Exinite + Vitrinite	30.4
% Vitrinite	26.3
% Sulfur (Total)	21.6
% Nitrogen	9.5
% Mineral Matter	1.5

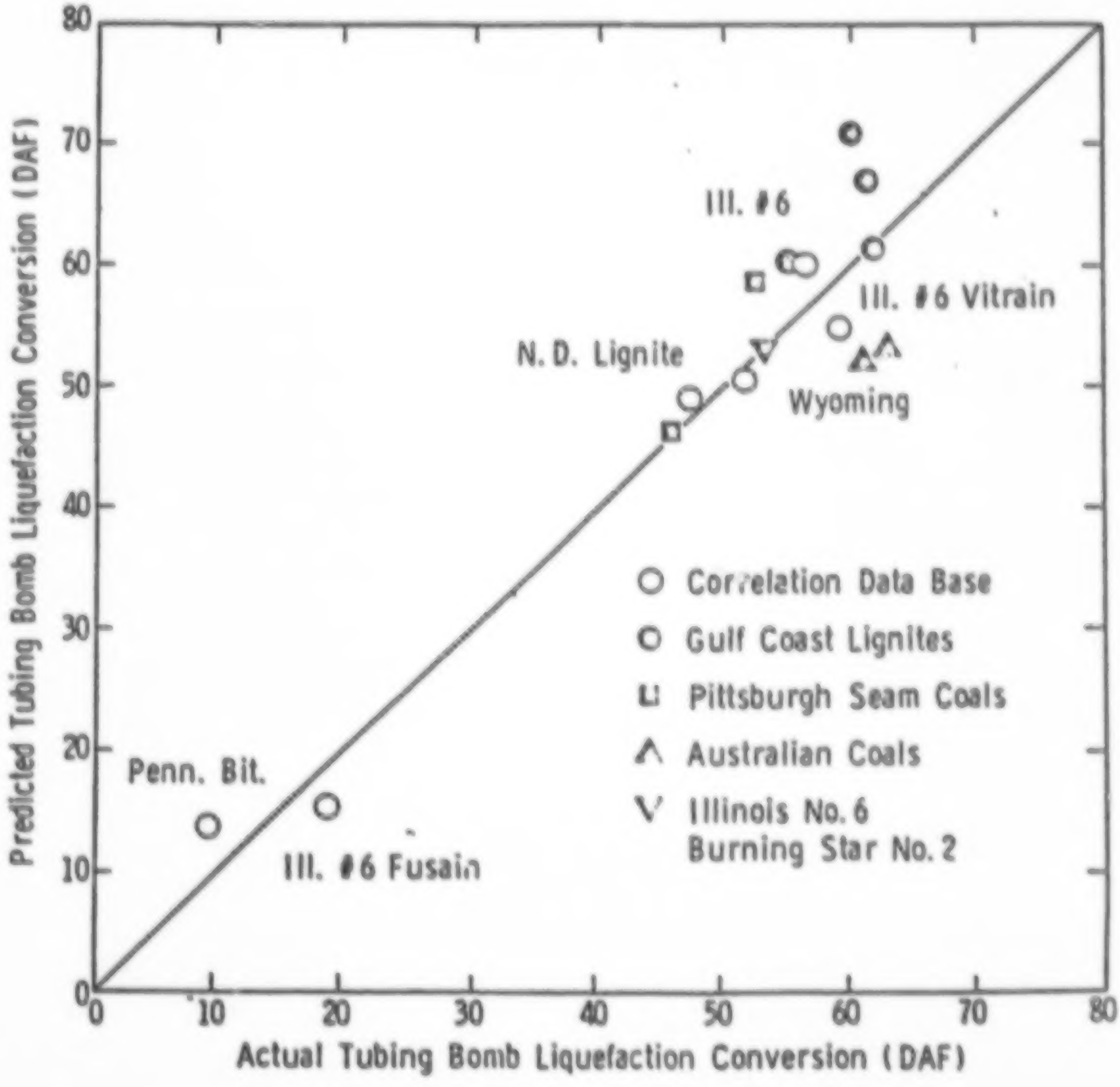
TABLE 4-4

SAMPLES ACQUIRED FOR SCREENING PROGRAM

Region	Seam	County	State	ASTM Rank
Rocky Mountain	08	San Juan	NM	SBA-HVC
	Wadge	Routt	CO	HVC
	Lower Sunnyside	Carbon	UT	HVB
Western Interior	Bevier	Howard	MO	HVC-HVB
	Fleming	Bourbon	KS	HVC-HVB
	Unnamed	Jasper	IO	HVC-HVB
Appalachian	Lower Kitanning	Cambria	PA	LV
	Pittsburgh	Somerset	PA	MV-HVA
	Pittsburgh	Washington	PA	HVA
	Upper Freeport	Beaver	PA	HVA

FIGURE 4-3

PARITY PLOT FOR TWO PARAMETER CORRELATION



HRI Microautoclave Tests

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MICROAUTOCLAVE TESTING

DETERMINATION OF POSSIBLE PRODUCT YIELDS FOR WESTERN CANADIAN COAL LIQUEFACTION CONSORTIUM

REPORT NO. L-12-64-(100)-501

MARCH 1982

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EXECUTIVE SUMMARY

The work described in this report was performed in accordance with the correspondence between Daniel M. Kennedy of Hydrocarbon Research, Inc. (HRI) and Eric D. Jamieson of the Western Canadian Coal Liquefaction Consortium, dated January 5, 1982. The work involved the evaluation of two British Columbian coals for use in the H-Coal™ Process using the microautoclave apparatus at HRI.

The objective of the program was to evaluate the Green Hills coal sample as a candidate for H-Coal Process operations by determining whether a set of operating conditions could be found for the microautoclave apparatus which would result in at least 40% distillate product yield. Additionally, tie lines were to be established with which HRI and the Western Canadian Coal Liquefaction Consortium could relate the results obtained from Green Hills coal to approximations of the behavior and expected yields from Monkman coal in an H-Coal Process system.

HRI projects that a 90% overall conversion could be achieved using Green Hills coal in H-Coal Process operations with resid recycle. Distillate yields of approximately 43% on a dry, ash-free charge basis are predicted.

Based on results obtained, HRI recommends further testing on the bench-scale and PDU level. Additional microautoclave tests may also be considered.

TECHNICAL PROGRAM OVERVIEW

This report describes the completed work required by the agreement between the Western Canadian Coal Liquefaction Consortium and Hydrocarbon Research, Inc. Under this contract, Green Hills coal was to be tested in HRI's microautoclave apparatus in order to determine probable yield distributions following treatment with the H-Coal Process. Monkman coal was to be evaluated relative to Green Hills for performance in the process as well.

Two British Columbian coal samples were received on January 21, 1982. The coals were prepared for microautoclave testing, and aliquots of the prepared samples were characterized. Western Canadian Coal Liquefaction Consortium can use the analytical data provided herein to estimate whether the samples sent are representative of the average sean analyses. Based on the reported properties, HRI has classified both Green Hills and Monkman coals as High-Volatile "A" (HVA) according to a modified ASTM ranking procedure.

The coals were tested in the microautoclave apparatus, which has two, 30-cc batch reactors. The microautoclave is useful in providing rough assessments of coal reactivities on a relative basis. Although not typically used as a method for determining yields or product distributions, HRI has developed a program which it feels can estimate this data reasonably well. It is emphasized that the results of this program cannot be used as a specific prediction of performance in a continuous-mode, H-Coal ebullated-bed reactor, but only as a general approximation.

Both coals were difficult to convert due to high inerts contents. However, data indicates that operation in a continuous H-Coal system with bottoms recycle should result in conversions of approximately 90% for Green Hills coal and somewhat lower for Monkman. Distillate yields for Green Hills should account for approximately 40-45% of the total product make.

HRI was unable to gather data concerning the effect of H_2S treatment on catalyst activity.

CONCLUSIONS AND RECOMMENDATIONS

CONCLUSIONS

Based on the acquired data and a knowledge of coal chemistry, HRI concludes the following:

- Green Hills coal can be processed effectively in an H-Coal system. Conversions of at least 90% should be attainable in an H-Coal Process system with vacuum bottoms recycle.
- A total distillate yield of approximately 43% was calculated for a steady-state process operating at conditions to yield 90% conversion of dry, ash-free coal to THF-soluble material. Middle and heavy distillates account for the greater part of this product group.
- Monkman coal can be expected to react in a manner similar to that of Green Hills. It did, however, have a slightly lower reactivity (approximately 7% lower) at high-severity conditions.

RECOMMENDATIONS

Based on microautoclave results and the H-Coal data base, the following ebullated-bed, bench unit program is proposed.

- Task 1 - Bench Unit Process Variables Study

  The ebullated-bed, bench unit shall be operated for a period of twenty-five days to establish preferred conditions for conversion of Western Canadian coals to distillate. Processing temperatures, pressures, residence times and compositions shall be explored during this experiment.

- Task 2 - Optimized Bench Run

  An ebullated-bed, bench unit run of thirty days' duration shall be conducted at the best set of operating conditions as determined from the process variable study. This operation will determine the catalyst deactivation pattern and projected catalyst usage. It will also determine the expected yields and conditions for operation on a larger scale and can serve as the basis for plant design.

- Task 3 - Process Development Unit Evaluation

  A 30-day, lined-out operation using the 3-1/2 ton/day Process Development Unit shall be conducted at conditions chosen for plant design. Catalyst shall be added and removed daily to achieve operation at equilibrated reactor catalyst conditions, product yields and recycle composition.

  Scaleup and larger-scale equipment performance shall be demonstrated. Sufficient product will be collected for upgrading studies and further testing.

Further microautoclave studies are recommended to investigate in more detail the two-stage, microautoclave conversion of Western Canadian coals.

HRI recommends that coal cleaning be investigated as a mechanism to improve the performance of B.C. coals in H-Coal operations.

A modified "Canadian oil agglomeration" technique is suggested for this investigation. The cleaning technique could be tested at HRI. The microautoclave apparatus could be used to identify increases in yields as a result of cleaning operations.

TECHNICAL PROCEDURES - MECHANICAL, PHYSICAL, ANALYTICAL

SAMPLE PREPARATION

Two British Columbian coal samples were received on January 21, 1982. These samples, Green Hills and Monkman, were stage-crushed under an inert atmosphere and split into 5-gram aliquots. The aliquots were placed in 8-gram sample bottles, which were then flushed with nitrogen and sealed with polethylene-lined caps. The caps were taped to prevent accidental opening and exposure to air. Aliquoting in this manner prevents excess oxidation caused by constantly opening and closing jars each time a sample is desired.

MICROAUTOCLAVE APPARATUS TEST PROCEDURE

The microautoclave is designed to rapidly heat and quench test charges while also providing vigorous agitation. Two, 30-cc reactors are operated simultaneously under hydrogen pressure, usually 2000 psig. A heated, fluidized sand bath provides heating, while rapid cooling is effected by agitating the reactor bombs in a water bath.

The reactor bombs are loaded with known quantities of coal and solvent, as well as catalyst, if called for. The bombs are then sealed and pressurized prior to being plunged into the sand bath, which has been preheated to a specified temperature. After waiting for the bombs to reach desired temperature, the bombs are agitated in the sand bath for a specified time period. Tests have been run to measure the required heating times. It has been found that a 2.5-minute interval is sufficient to allow the bombs to heat to a temperature of 850°F. Once the bombs have been held at temperature for the required time, they are removed from the sand bath and plunged into a water bath to be cooled. After cooling, the bombs are depressurized and weighed. The material inside is removed and analyzed. Analysis consists of either a THF extraction or a toluene-THF series extraction followed by an ash determination.

Conversions are calculated by subtracting the organic product insolubles from the total organic feed insolubles (both coal and recycle resid) and dividing that quantity by the coal ash-free weight. Multiplication by 100 gives a value of percent conversion that is meaningful for both a once-through process and a steady-state recycle process approximation.

All tests in this program used hydrogen at a pressure of 2000 psig. Temperatures studied ranged from 750 to 850°F, and holding times were either 10, 15, or 30 minutes (12-1/2, 17-1/2, and 32-1/2 minutes total run time, including heatup and cooldown).

The program itself had several segments. In the first segment, both Green Hills and Monkman coals were each subjected to ten tests at various temperatures and times without using any catalyst. Based on these tests, an optimum set of conditions was determined for Green Hills first-stage treatment, and data was gathered to try to establish tie lines between Green Hills and Monkman conversion behavior. Once a set of conditions was chosen for Green Hills thermal conversion, a four-point, two-stage test series was conducted. In this series, the first stage involved purely thermal treatment at conditions chosen based on results from the first portion of the program. The second stage was run in catalytic mode and was run at four time-temperature conditions.

Two-stage microautoclave tests were conducted in the following manner. The bomb was charged in the normal manner and run without catalyst for a specified time at a specified temperature. The bomb was then cooled and depressurized and opened. The products remained inside and a basket containing a certain amount of catalyst is inserted to the bomb. The bomb is closed, repressurized, and run at another set of sepcified conditions. It is then cooled, depressurized, and analyzed.

Based on the results obtained in this second phase of testing, a set of conditions was chosen for two-stage operations to produce a larger quantity of products (about 150 g) for a more detailed analysis, as well as for use as solvent-recycle material for the next phase of the program. In this way, the recycle process used by a continuous feed reactor could be approximated. This two-pass, two-stage operation was conducted as follows. Material produced in the first pass procedure is loaded into a bomb with some fresh coal. This sample is run thermally and then in catalytic mode according to the same procedure outlined in the previous paragraph. Products are then removed and analyzed.

In all first pass operations, the solvent was hydrogenated anthracene oil and the bombs were loaded with a solvent-to-coal ratio of 2. Second pass operations used solvent to coal ratios of 8. This is due to the fact that the solvent, products of the first pass procedure, contain enough residual material to make the solvent to solids ratio close to 2, which is a desirable target. Catalyst charges were always 0.5 g catalyst per gram of solids.

KINETIC MODEL

Because a mixture of time and temperature conditions are explored simultaneously, a model is needed to normalize the conditions to one basis, a Standard Temperature Time Unit (STTU). HRI has found that liquefaction can be modeled by an assumed first-order reaction. Using this model, the STTU's for a particular set of conditions can be calculated. This provides a means of determining the relative severities of different sets of operating conditions.

SOLVENTS

The solvent used in all first pass tests was hydrogenated anthracene oil. Typical properties of this solvent are listed in Table 1.

The solvent used in second pass operations was the total product of the first pass procedure. Analysis indicates that this material had an initial boiling point of approximately 425°F, and contained 31% unconverted solids. The use of such a material simulates both vacuum bottoms recycle, and the use of middle to heavy distillate products to slurry the fresh coal feed.

COAL PROPERTIES

Detailed proximate and ultimate analyses were performed on both Green Hills and Monkman coals by HRI. Table 2 contains a partial list of the property data collected on the coals (see Appendix A for the complete analyses).

ASTM-ranking procedures gave ambiguous results when applied to either coal sample. The volatile matter content, used as the ranking parameter, suggests that both coals are medium-volatile. Using the calculated calorific values, however, as the ranking factor dictates that both coals are high-volatile "B" bituminous (HVB) rank. Mean reflectance indicates that both coals are low-volatile. HRI feels that these discrepancies are due to the relatively high inertinite contents of both coals. This high content of inerts would tend to suppress the volatile content and the calorific values, while increasing the mean reflectance of the overall coal samples. If the reported coal data is used to generate volatile matter content and calorific values on a dry, mineral matter, inerts-free basis, then both coals would be ranked as high-volatile "A" (HVA) coals. HRI feels that this is a valid ranking since a strict ASTM-ranking procedure is generally invalid for high inert coals such as Green Hills or Monkman.

The high inert content is evident from first pass operations, as once-through conversions are limited to 70-75% regardless of reactor conditions.

TABLE 1

TYPICAL ANALYSES OF HYDROGENATED ANTHRACENE OIL EMPLOYED IN MICROAUTOCLAVE TESTS

HRI HAO (HRI No. L-532[4])

Gravity, °API	-0.8
IBP, °F	476
10, V %	524
30, V %	571
50, V %	610
70, V %	650
90, V %	727
EP, V %	831
V % at 650°F	70
V % at EP	97
W % at 650°F	67
W % at EP	89
W % at EP$^+$	4.1
C, %	90.60
H, %	7.58
N, %	0.28
S, %	0.15

TABLE 2

SUMMARY OF COAL PROPERTIES

	GREEN HILLS	MONKMAN
PROXIMATE ANALYSIS		
MOISTURE, AS RECEIVED , WT%	1.48	.98
ASH, DRY, WT%	14.35	16.58
VOLATILES, DMMF , WT%	22.93	20.72
FIXED CARBON, DMMF, WT%	77.07	79.28
ULTIMATE ANALYSIS		
CARBON, DMMF, WT%	90.51	92.18
HYDROGEN, DMMF, WT%	4.71	4.79
NITROGEN, DMMF, WT%	1.23	1.01
OXYGEN, DMMF, WT%	3.17	1.59
TOTAL SULFUR, DMMF, WT%	.37	.43
TOTAL SULFUR, DRY	.45	.46
H/C RATIO, DAF,	.63	.62
O/C RATIO, DAF,	.04	.03
PETROGRAPHY		
VITRINITE CONTENT, DMMF , VOL%		
VITRINITE MEAN REFLECTANCE, %		
QUINOLINE INSOLUBLES, WT%		
THF INSOLUBLES, WT%	95.98	95.90
TOLUENE INSOLUBLES, WT%	97.01	97.50
FREE SWELLING INDEX ,	3.0	3.0

RESULTS

GREEN HILLS TEST PROGRAM

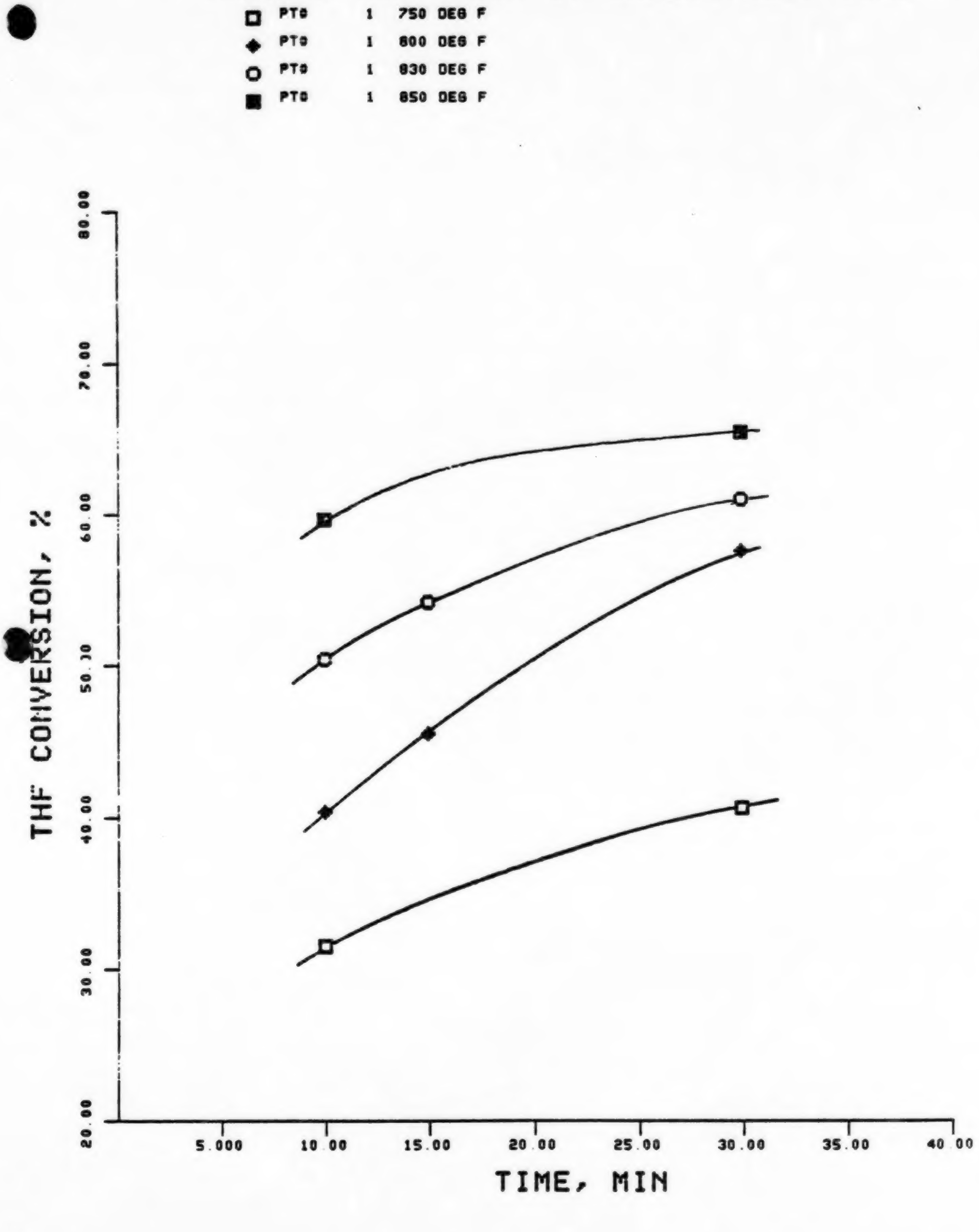
The results obtained from the first phase of the project conducted on Green Hills coal are found in Figure 1. This plot indicates that conversion to THF-soluble material increases with both time and temperature over the ranges studied. This information was used to choose a suitable set of conditions for the first stage of a two-stage procedure to be used in the next portion of the program. The plan was to choose a set of first-stage conditions which provided a high degree of conversion without decreasing solvent quality excessively. For these reasons, 850°F and ten minutes were chosen as the temperature and holding time to be used in the first stage. These conditions yield about 60% conversion of coal to THF-soluble material with hydrogenated anthracene oil as solvent.

The next portion of the program involved a two-stage operation, with the first stage a purely thermal treatment, while the second involved catalytic treatment of the Green Hills coal. The data obtained from the four tests that were run are found in Figure 2. Both toluene and THF extractions were performed on the products obtained from these tests in order to determine the conversion to 975°F$^-$ liquids, as well as the overall conversion. Each of the four tests was conducted using the conditions set forth above in the first, or thermal, stage, while conditions in the catalytic second stage were varied. The results indicated that maximum conversion could be achieved by operating the second stage at 850°F with a holding time of ten minutes. Operation at these conditions provided yields of 72% THF conversion and 67% toluene conversion with hydrogenated anthracene oil as the solvent.

The third phase of the program attempted to simulate the continuous H-Coal Process more closely by the introduction of recycle. Since the continuous operation process calls for the use of distillates as solvent and the recycling of unconverted solids, it was decided that, in the next phase, hydrogenated anthracene would not be used as the coal solvent. Instead, the products from a once-through, two-stage microautoclave run would be used.

FIGURE 1

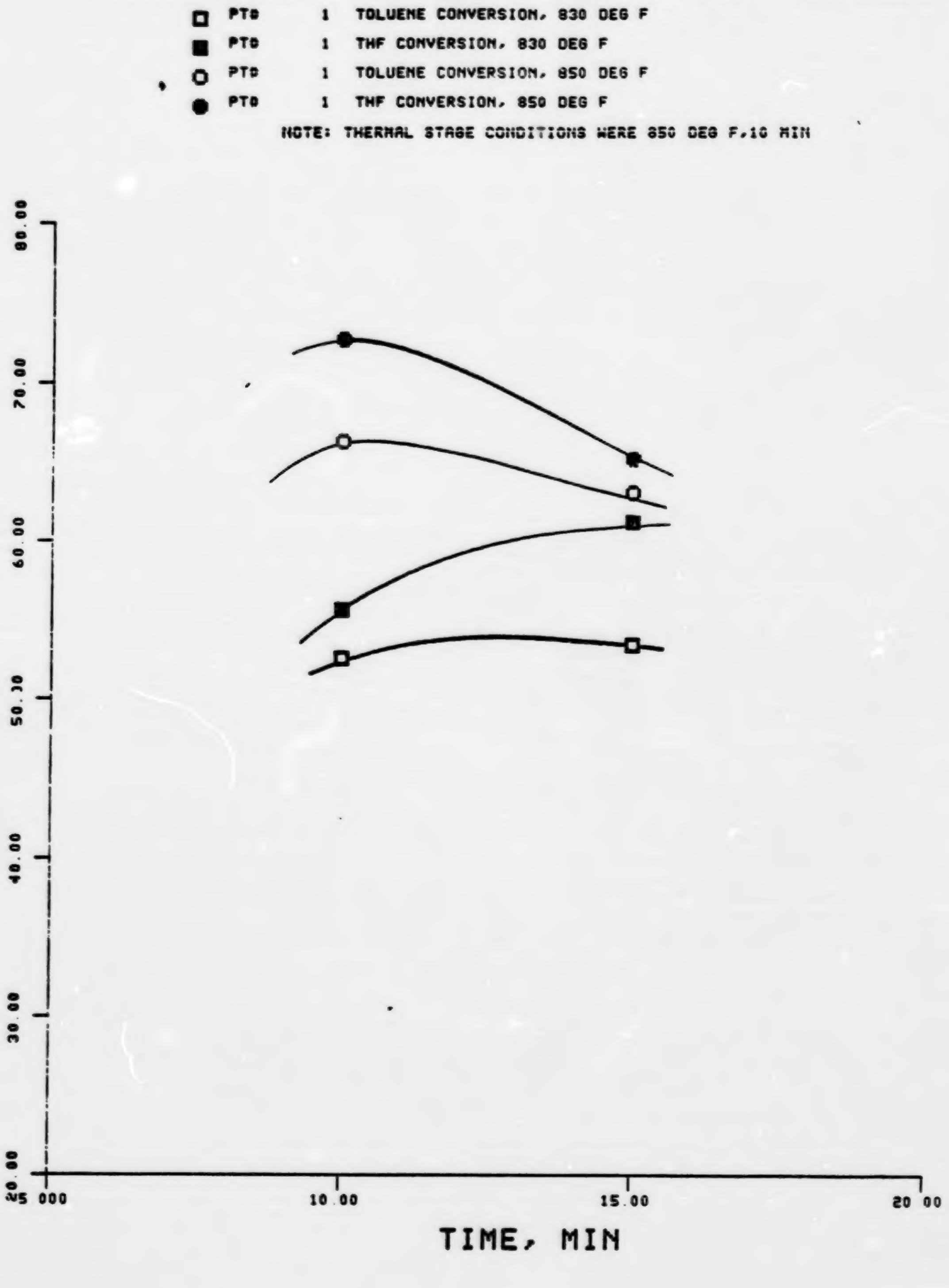
THERMAL TREATMENT OF GREEN HILLS COAL



2.

FIGURE 2

SIMULATION OF ONCE-THROUGH H-COAL PROCESS



Thermal-stage conditions were fixed at 850°F for ten minutes, while four sets of catalytic stage conditions were tried. The resulting data was difficult to interpret due to product losses and poor ash balances, but it was determined that operation of the catalytic stage at 850°F for fifteen minutes yielded optimum results of 82.14% toluene conversion and 105.09% THF conversion.

These results cannot be used as estimates of steady-state process yields, but they do indicate that recycled solids are converted, as well as the fresh coal feed.

Based on previous experience on the bench-scale, HRI feels that it should be possible to operate a continuous, steady-state process at the 90% THF conversion level. If the toluene extraction data is scaled down accordingly, conversion of dry, ash-free coal to toluene solubles is predicted to be approximately 70.4%.

A large quantity of material was produced in the microautoclave running at the optimum conditions set forth above for the two-stage, two-pass operation. This material was analyzed at HRI to provide the necessary product distribution and distillate yield data. In order to determine the true coal product yield, it must be remembered that a portion of the product liquid is hydrogenated anthracene and not coal-derived liquid. Since the hydrogenated anthracene itself was produced at conditions similar to those used in these tests, it is assumed that none of the solvent is affected by the process. Some of the hydrogenated anthracene will, however, vaporize and be lost from the bomb during depressurization, as will coal-derived gases and light distillates. As the microautoclave has no provision to recover these lost materials, a direct, quantitative determination of product yield distribution is very difficult. An indirect determination, however, can be readily performed.

Figure 3 shows the total recovered product yield distribution as given by an ASTM D-1160 vacuum distillation. Figure 4 shows a boiling point distribution curve for hydrogenated anthracene. If an arbitrary value is chosen for hydrogenated anthracene content in the total product, the curve in Figure 4 can be rescaled and subtracted from the curve in Figure 3 to give the curve seen in Figure 5. Note that the arbitrary choice of anthracene content will not affect the shape of the difference curve, except in the region between approximately 425 and 475°F where there should be no anthracene.

FIGURE 3

B.P. DISTRIBUTION OF TOTAL COLLECTED PRODUCT

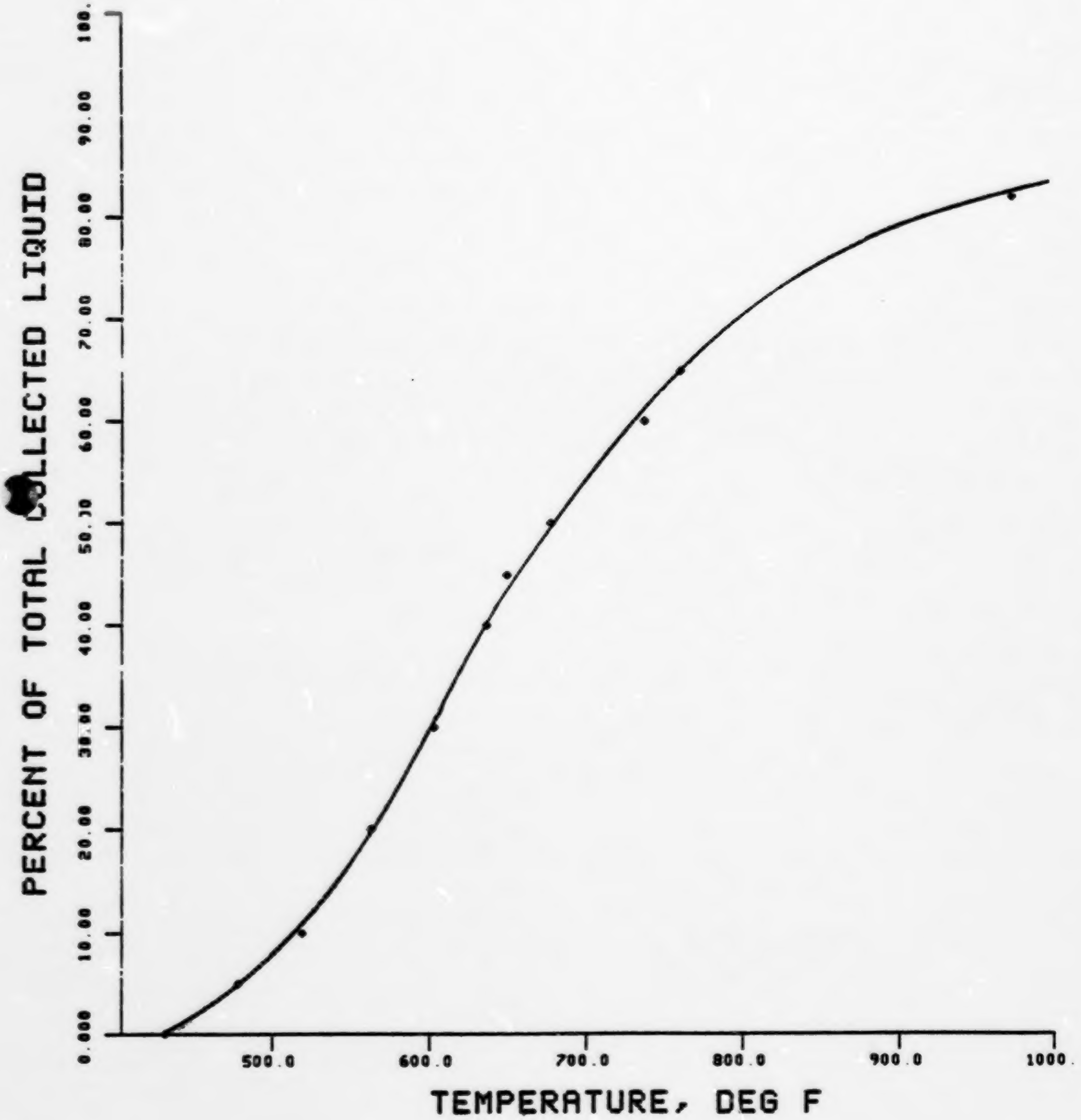


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FIGURE 4

B.P. DISTRIBUTION OF HYDROGENATED ANTHRACENE OIL

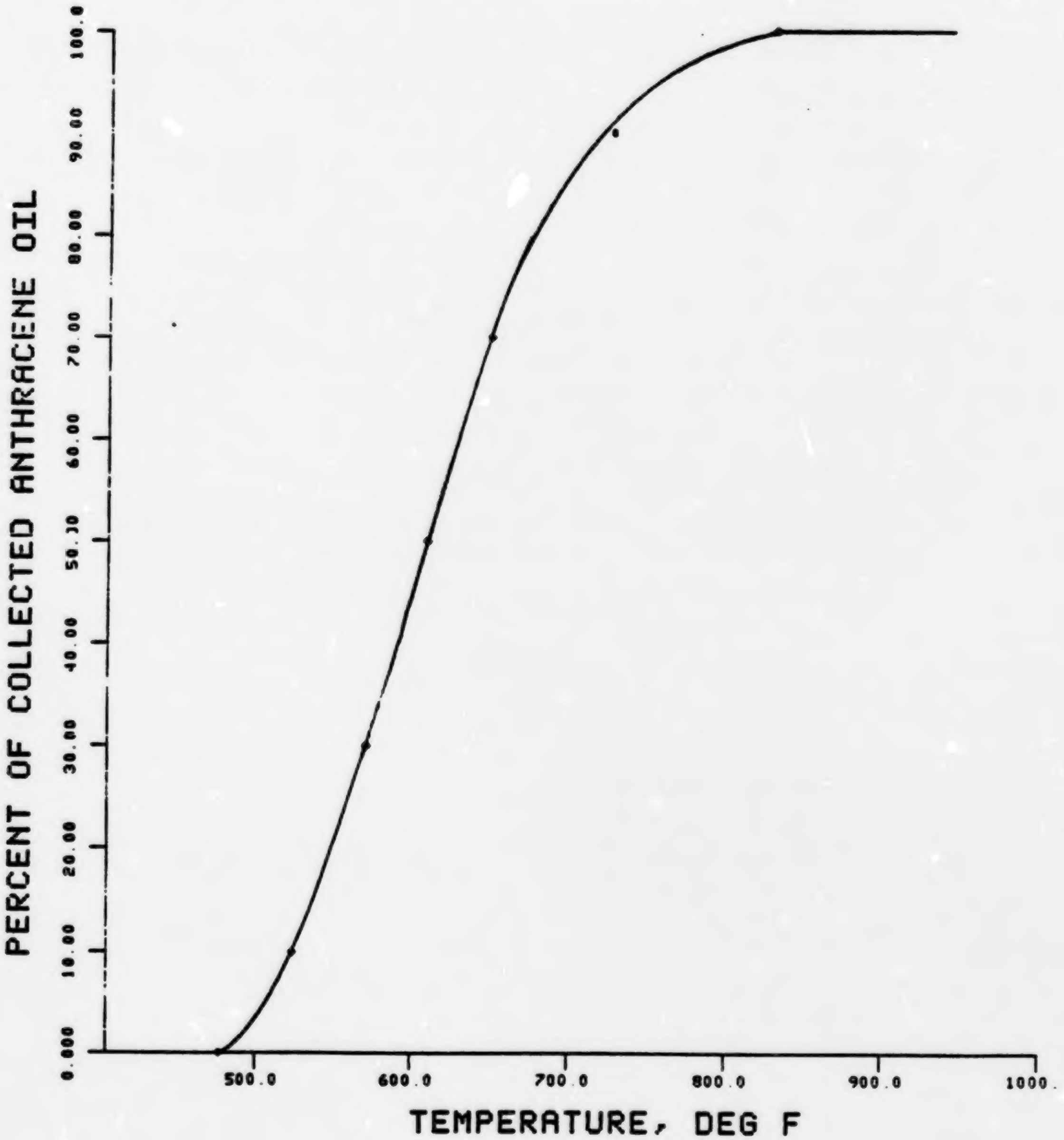
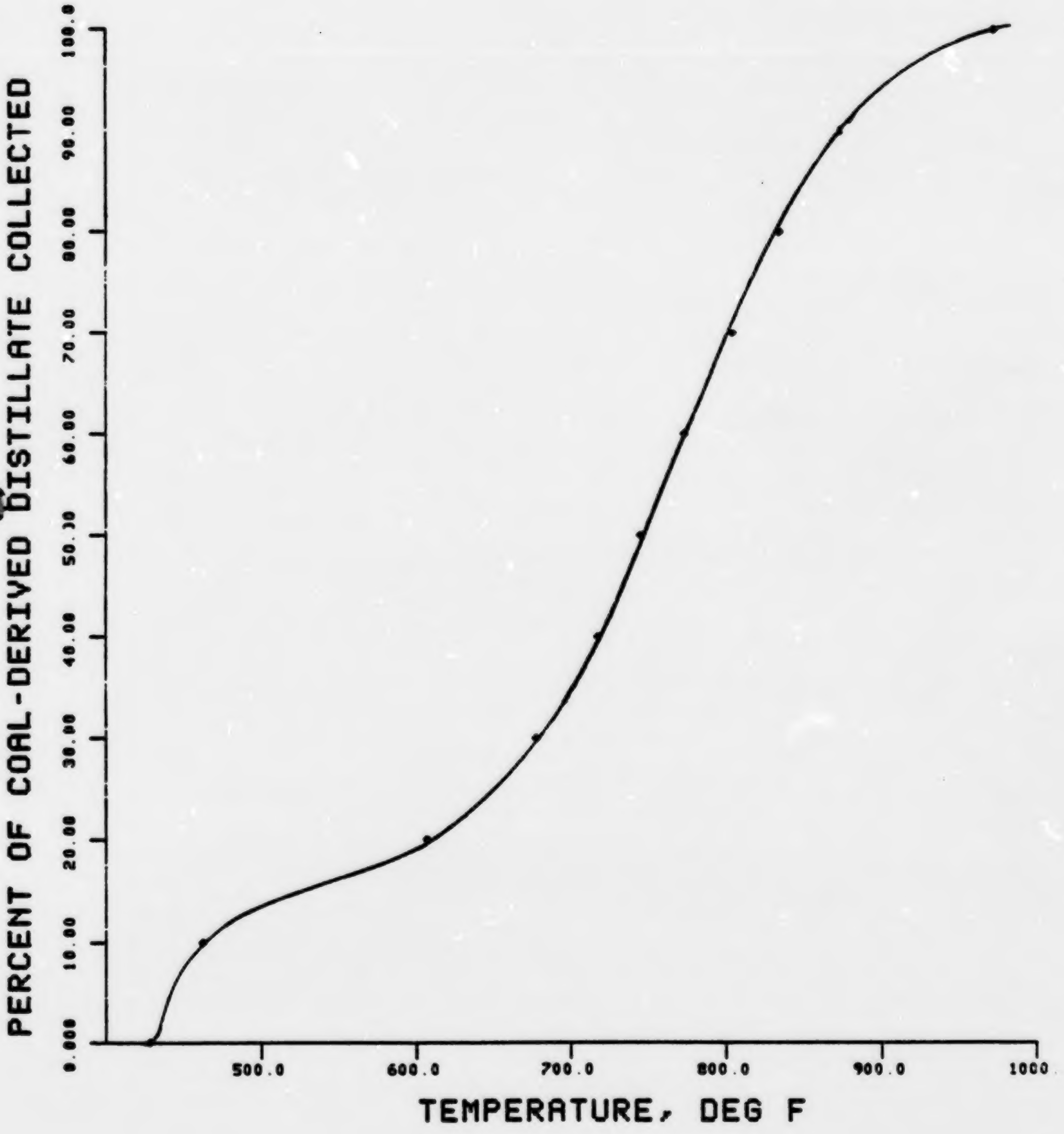


FIGURE 5

B.P. DISTRIBUTION OF COAL - DERIVED PRODUCT LIQUIDS



An estimate of 50% hydrogenated anthracene content was used to generate the specific curve in Figure 5. Some data was obtained to indicate that the actual content should be between 40 and 60%. Since the product fraction between 425 and 475°F is small, variations in calculated yields due to improper estimation of hydrogenated anthracene content should also be small. A more critical assumption in the generation of the curve in Figure 5 is that any hydrogenated anthracene loss is distributed evenly across the entire boiling range of the anthracene. Although this assumption is not likely to be valid, it will serve to provide a worse case estimate of product yield distributions.

The full range of product distributions is not taken directly from Figure 5. Note that this curve is caled to provide percentages of coal-derived distillate products only. By using the toluene and THF extraction data, the actual distillate quantity can be determined for the projected steady-stage case. Prior work with coals having compositions similar to that of Green Hills coal indicates that approximately 75% of the toluene-soluble material should have boiling points less than 975°F. THF extraction results are used to directly obtain the total coal-derived liquid make. Thus, when operating at conditions to provide a 90% THF conversion, 52.8% of the ash-free coal charge should be converted to material which boils below 975°F.

Previous experience with similar coals in bench- and PDU-scale operations indicates that the total gas production should be approximately 9.5 to 10.0% of the ash-free coal charge. This gas consists of hydrocarbons with one to six carbon atoms, as well as H_2O, NH_3 and H_2S. Assuming that all of the oxygen, nitrogen, and sulfur in the converted coal is hydrogenated, the chemical gas yield will be 4.9%. Using the above information with the curve in Figure 5, the following product yields are estimated:

Chemical Gases (H_2O, NH_3, H_2S)	4.9%
Hydrocarbon Gases (C_1-C_6)	4.9
Light Distillate (IBP-400°F)	1.8
Middle Distillate (400-650°F)	5.8
Heavy Distillate (650-975°F)	35.4
Residual Liquid (975°F$^+$)	37.2
Unconverted Coal	10.0
Total	100.0%

The distribution within the distillate fractions is artifically shifted to the heavy end due to the assumption that hydrogenated anthracene losses were evenly distributed over the entire boiling point range of the material. If it is assumed that the hydrogenated anthracene losses are not evenly distributed, then the shape of the curve in Figure 4 changes. Consequently, the distribution in Figure 5 changes as well. As an example, a shift of 75°F in the initial boiling point of the hydrogenated anthracene oil in the product would result in coal liquid yields that could be approximated as follows:

Chemical Gas (H_2O, NH_3, H_2S)	4.9
Hydrogen Gas (C_1 to C_4)	4.9
Light Distillate (IBP-400°F)	3.7
Middle Distillate (400-650°F)	19.9
Heavy Distillate (650-975°F)	19.4
Residual Liquid (975°F$^+$)	37.2
Unconverted Coal	10.0
Total	100.0

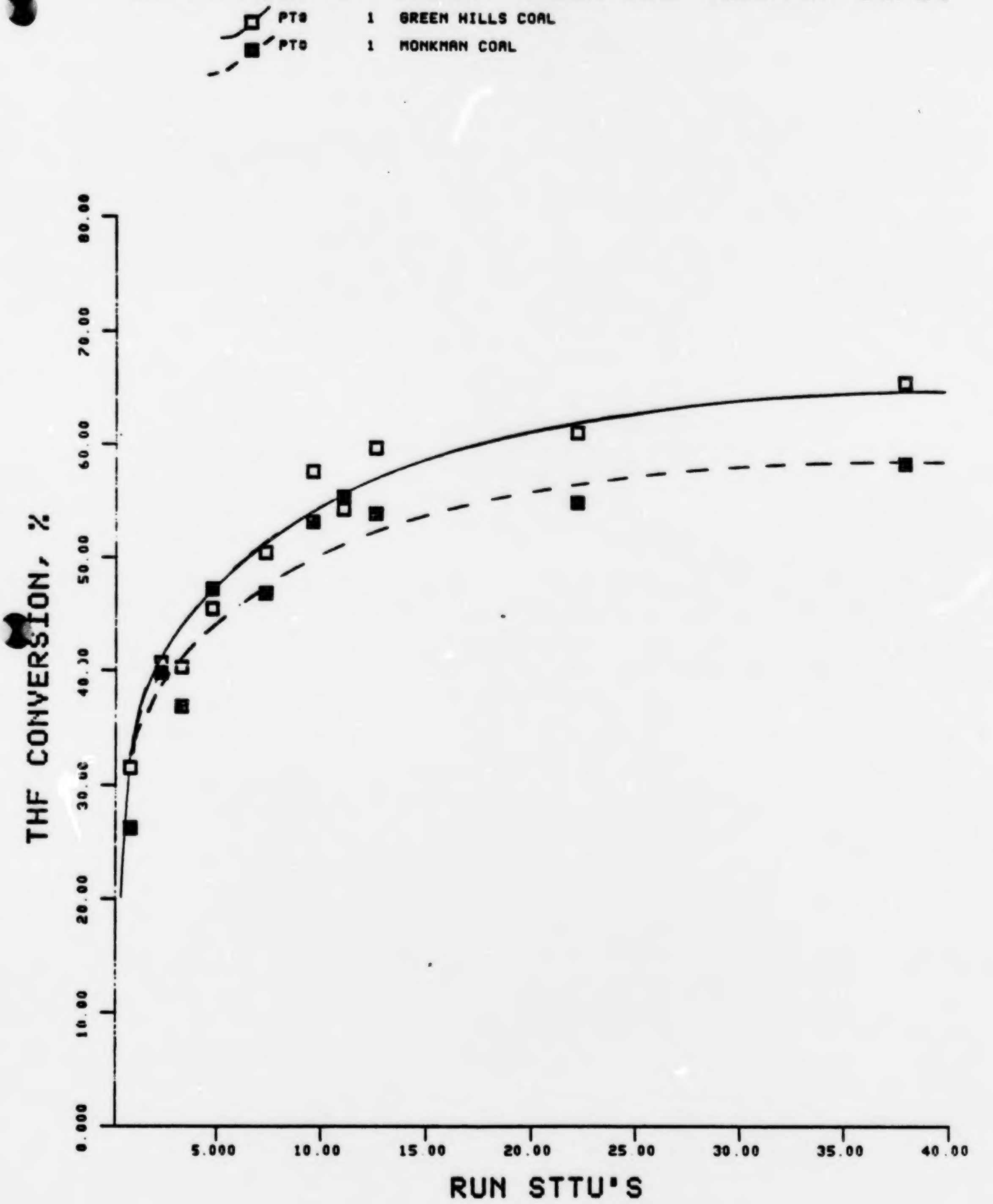
Although these yields are most likely more representative of the actual product distribution, further tests would have to be performed to determine more specifically the behavior of hydrogenated anthracene in the microautoclave apparatus. This information would permit a more accurate determination of product yields.

MONKMAN TEST PROGRAM

In order to establish tie lines between the behaviors of Green Hills and Monkman coals, a ten-test, thermal series was conducted on the Monkman coal. The results of these tests were compared to those from the first phase of the Green Hills program. Figure 6 shows the thermal conversions achieved with each coal as a function of Standard Temperature-Time Units (STTU's) which are a measure of the severity of liquefaction operating conditions. This plot shows that at low severities, less than 5 STUU's, both coals might be expected to behave in a similar manner in a liquefaction process. At higher severities, however, Monkman coal exhibits a lower degree of conversion than does the Green Hills. At 850°F with a thirty-minute hold time (38.09 STTU's), this amounts to a 7% lower degree of conversion. It is important to note that neither coal shows occurrance of re-condensation reactions at high severities, which would be indicated by a decreasing conversion with increasing STUU's at high severity.

FIGURE 6

COMPARISON OF GREEN HILLS AND MONKMAN COALS



APPENDIX A

DETAILED COAL ANALYSES

TABLE A-1

GREEN HILLS HRI # 4465

COMPONENT	AS RECEIVED	DRY	DAF	DMMF
MOISTURE	1.480			
ASH	14.138	14.350		
VOLATILES	20.591	20.900	24.402	22.935
FIXED CARBON	63.792	64.750	75.598	77.065
TOTAL	100.000	100.000	100.000	100.000
CARBON	74.960	76.086	88.712	90.514
HYDROGEN	4.200	4.095	4.617	4.711
NITROGEN	1.020	1.035	1.209	1.233
SULFUR	.443	.450	.478	.369
OXYGEN	12.245	11.095	4.984	3.172
PHOSPHORUS	.038	.039		
SILICON	3.749	3.805		
ALUMINUM	1.966	1.995		
IRON	.872	.885		
CALCIUM	.151	.153		
MAGNESIUM	.056	.057		
SODIUM	.023	.023		
POTASSIUM	.158	.161		
TITANIUM	.119	.120		
STRONTIUM	0.000	0.000		
BARIUM	0.000	0.000		
MANGANESE	0.000	0.000		
TOTAL METALS	7.239	7.132		
TOTAL	100.000	100.000	100.000	100.000
H:C RATIO	.673	.646	.625	.625
O:C RATIO	.123	.109	.042	.026
MINERAL MATTER, WT %			16.06	

MEASURED CALORIFIC VALUE, DRY	
CORRECTED CALORIFIC VALUE, DRY	
CALCULATED IGT CALORIFIC VALUE (DRY BASIS)	12775.60
CALCULATED MS CALORIFIC VALUE (DMMF BASIS)	13241.82
CALCULATED EPE CALORIFIC VALUE (DRY,ORGANIC)	11542.97

TABLE A-2

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MONKMAN                          HRI # 4466

COMPONENT      AS RECEIVED         DRY              DAF              DMMF

MOISTURE              .980
ASH                 16.418        16.580
VOLATILES           18.477        18.660           22.369           20.717
FIXED CARBON        64.125        64.760           77.631           79.283
               ----------     ----------       ----------       ----------
TOTAL              100.000       100.000          100.000          100.000

CARBON              74.880        75.621           90.471           92.178
HYDROGEN             4.130         4.060            4.704            4.792
NITROGEN              .820          .828             .993            1.011
SULFUR                .455          .460             .397             .427
OXYGEN              11.659        10.896            3.436            1.591

PHOSPHORUS            .026          .026
SILICON              5.015         5.065
ALUMINUM             1.800         1.817
IRON                  .454          .458
CALCIUM               .218          .220
MAGNESIUM             .111          .112
SODIUM                .076          .076
POTASSIUM             .259          .261
TITANIUM              .097          .098
STRONTIUM            0.000         0.000
BARIUM               0.000         0.000
MANGANESE            0.000         0.000
               ----------     ----------
TOTAL METALS         8.135         8.055


               ----------     ----------       ----------       ----------
TOTAL              100.000       100.000          100.000          100.000

H:C RATIO             .662          .645             .624             .624
O:C RATIO             .117          .108             .029             .013

MINERAL MATTER, WT %                               18.12

MEASURED CALORIFIC VALUE, DRY
CORRECTED CALORIFIC VALUE, DRY
CALCULATED IGT CALORIFIC VALUE (DRY BASIS)      12694.16
CALCULATED MS CALORIFIC VALUE (DMMF BASIS)      13235.70
CALCULATED ERE CALORIFIC VALUE (DRY,ORGANIC)    11531.34
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APPENDIX B

SUMMARY OF MICROAUTOCLAVE TEST DATA

TABLE B-1

ONCE-THROUGH THERMAL TREATMENT OF GREEN HILLS COAL

Solvent: Hydrogenated Anthracene

Coal Charged, g	2.4984	2.4948	2.4997	2.5042	2.5048
Solvent Charged, g	5.0	5.0	5.0	5.0	5.0
Temperature, °F	750	750	800	800	800
Time, Min.	10	30	10	15	30
Ash-Free Weight of Charge, g	2.1399	2.1368	2.1410	2.1448	2.1454
Total THF-Insolubles in Produce, g	1.8012	1.6270	1.6335	1.5303	1.2432
Ash in Product, g	0.3379	0.3608	0.3592	0.3637	0.3351
Organic THF-Insoluble in Product	1.4633	1.2662	1.2743	1.1666	0.9081
THF Conversion, %	31.62	40.74	40.48	45.61	57.67

TABLE B-1
(CONTINUED)

ONCE-THROUGH THERMAL TREATMENT OF GREEN HILLS COAL (CONTINUED)

Solvent: Hydrogenated Anthracene

Coal Charged, g	2.5132	2.4929	2.4900	2.4943	2.4986
Solvent Charged, g	5.0	5.0	5.0	5.0	5.0
Temperature, °F	830	830	830	850	850
Time, Min.	10	15	30	10	30
Ash-Free Weight of Charge, g	**2.1526**	**2.1352**	**2.1327**	**2.1364**	**2.1401**
Total THF-Insolubles in Produce, g	**1.4065**	**1.3569**	**1.1920**	**1.2039**	**1.0672**
Ash in Product, g	**0.3407**	**0.3815**	**0.3637**	**0.3433**	**0.3291**
Organic THF-Insoluble in Product	**1.0658**	**0.9754**	**0.8283**	**0.8606**	**0.7381**
THF Conversion, %	**50.49**	**54.32**	**61.16**	**59.72**	**65.51**

TABLE B-2

SIMULATION OF ONCE-THROUGH H-COAL PROCESS

Solvent: Hydrogenated Anthracene

Coal Charged, g	2.0042	2.0015	1.9962	1.9941
Solvent Charged, g	4.0	4.0	4.0	4.0
First Stage				
Temperature, °F	850	850	850	850
Time, Min	10	10	10	10
Second Stage				
Temperature, °F	830	830	850	850
Time, Min.	10	15	10	15
Catalyst Added, g	1.2610	1.4234	1.3854	1.4200
Ash-Free Weight of Charge, g	1.7166	1.7143	1.7097	1.7079
Toluene-Insoluble in Product, g	1.0856	1.2366	0.8067	0.9203
THF-Insoluble in Product, g	1.0335	1.1036	0.6955	0.8840
Ash in Product, g	0.2720	0.4394	0.2301	0.2899
Organic Toluene-Insoluble in Product, g	0.8136	0.7972	0.5766	0.6304
Organic THF-Insoluble in Product, g	0.7615	0.6642	0.4654	0.5941
Toluene Conversion, %	52.60	53.50	66.28	63.09
THF Conversion, %	55.64	61.26	72.78	65.22

TABLE B-3

SIMULATION OF H-COAL PROCESS WITH RESID RECYCLE

Solvent: First Pass - Hydrogenated Anthracene
Second Pass - Products from First Pass

Note: First Pass Conditions Were Set at 850°F, 10 Minutes for Both Thermal and Catalytic Stages.

Coal Charged to First Pass	2.9938
Coal Charged to Second Pass	0.9999
Solvent Charged to First Pass	6.0
Conditions of Second Pass	
Temperature Stage 1/Stage 2	850/850
Time Stage 1/Stage 2	10/15
Second Pass Solvent	
Calculated Org. Tol. Insol.	0.8648
Calculated Org. THF. Insol.	0.6980
Calculated Ash in Solvent	0.4296
Ash-free Weight of Second Coal Charge	0.8564
Organic Tol. Insol. in Second Pass Charge	1.7212
Organic THF Insol. in Second Pass Charge	1.5544
Tol. Insol. in Final Product	1.4455
THF Insol. in Final Product	1.0822
Ash in Final Product	0.4278
Organic Tol. Insol. in Product	1.0177
Organic THF Insol. in Product	0.6544
Tol. Conversion	82.14
THF Conversion	105.09

TABLE B-4

•ONCE-THROUGH THERMAL TREATMENT OF MONKMAN COAL

Solvent: Hydrogenated Anthracene

Coal Charged, g	2.5055	2.5024	2.5071	2.4975	2.5017
Solvent Charged, g	5.0	5.0	5.0	5.0	5.0
Temperature, °F	750	750	800	800	800
Time, Min.	10	30	10	15	30
Ash-Free Weight of Charge, g	2.0901	2.0875	2.0914	2.0834	2.0869
Total THF-Insolubles in Produce, g	1.9390	1.6357	1.7074	1.4800	1.3716
Ash in Product, g	0.3976	0.3822	0.3896	0.3838	0.3963
Organic THF-Insoluble in Product	1.5414	1.2535	1.3178	1.0962	0.9753
THF Conversion, %	26.25	39.95	36.99	47.38	53.27

TABLE B-4
(CONTINUED)

ONCE-THROUGH THERMAL TREATMENT OF MONKMAN COAL (CONTINUED)

Solvent: Hydrogenated Anthracene

Coal Charged, g	2.4988	2.5044	2.5006	2.4930	2.4997
Solvent Charged, g	5.0	5.0	5.0	5.0	5.0
Temperature, °F	830	830	830	850	850
Time, Min.	10	15	30	10	30
Ash-Free Weight of Charge, g	2.0845	2.0892	2.0860	2.0797	2.0852
Total THF-Insolubles in Produce, g	1.5009	1.2802	1.3361	1.3312	1.2543
Ash in Product, g	0.3959	0.3492	0.3972	0.3752	0.3878
Organic THF-Insoluble in Product	1.1050	0.9310	0.9389	0.9560	0.8665
THF Conversion, %	46.99	55.44	54.99	54.03	58.45